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A Study of Nitrate Respiration in the Activated Sludge Process

New York State Dept. of Environmental
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16. ABSTRACT

In an experimental, 570-m^3/day (0.15-mgd) activated sludge plant treating domestic wastewater from a correctional facility, 76 to 87 percent nitrogen removal was obtained via sludge synthesis and biological denitrification using endogenous H-donors in a compartmentalized reactor with alternating aerobic and anoxic zones. Between 27 and 48 percent of the influent nitrogen was removed by denitrification and between 37 and 49 percent via sludge synthesis. The process was operated for 8 mo under comprehensive analytical control. Ferric chloride ($FeCl_3$) was used to enhance phosphorus removal. The lowest winter temperature measured in the aeration tank was 15.9 C. An in-line surge tank was employed for flow equalization. Primary settling was not utilized in the first 5 mo of operation during the 1974-75 winter. A turbid effluent developed in the fifth month of operation, however. This condition was finally brought under control by adding primary settling to the flow scheme and dosing $FeCl_3$ to the influent of the primary settler. This change in treatment strategy reduced nitrogen removal from 82 percent to 78 percent by reducing sharply the amount of nitrogen removed via sludge synthesis. The portion of nitrogen removed by denitrification was not affected. This type of operation was used for 3 mo.

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EPA-600/2-80-154
August 1980

A STUDY OF NITRATE RESPIRATION IN THE ACTIVATED SLUDGE PROCESS

by

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10

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

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This report describes a successful effort to develop a modification of the activated sludge process that removes a high percentage of nitrogen from municipal sewage without the addition of an organic carbon source like methanol.

Francis T. Mayo, Director
Municipal Environmental Research
Laboratory

ABSTRACT

In an experimental, 570-m^3/day (0.15-mgd) activated sludge plant treating domestic wastewater from a correctional facility, 76 to 87 percent nitrogen removal was obtained via sludge synthesis and biological denitrification using endogenous H-donors. Between 27 and 48 percent of the influent nitrogen was removed by denitrification and between 37 and 49 percent via sludge synthesis. The process was operated for 8 mo under comprehensive analytical control. Ferric chloride ($FeCl_3$) was used to enhance phosphorus removal. The lowest winter temperature measured in the aeration tank was 15.9 C. An in-line surge tank was employed for flow equalization. Primary settling was not utilized in the first 5 mo of operation during the 1974-75 winter. A turbid effluent developed in the fifth month of operation, however. This condition was finally brought under control by adding primary settling to the flow scheme and dosing $FeCl_3$ to the influent of the primary settler. This change in treatment strategy reduced nitrogen removal from 82 percent to 78 percent by reducing sharply the amount of nitrogen removed via sludge synthesis. The portion of nitrogen removed by denitrification was not affected. This type of operation was used for 3 mo.

The surge tank was used as part of the activated sludge system for short periods of time. Pulsating aeration by a floating aerator controlled by a time clock in the surge tank allowed the direct use of wastewater carbon as the H-donor in biological denitrification. Nitrogen removal efficiency was thereby increased, but a bulking sludge problem developed.

By chlorinating excess activated sludge or a mixture of primary and excess activated sludge to pH 2.5 and using the pipeline from the chlorinator to the sludge drying bed as a reactor, a 50 percent increase in sludge filterability was obtained.

The feasibility of utilizing a chlorine contact chamber as a multiple tray second-phase settling tank featuring a cross-flow arrangement of settling surfaces was investigated. An incremental suspended solids removal of 33 percent was obtained.

This report was submitted in fulfillment of Grant No. 17050 EDL by the New York State Department of Environmental Conservation under the partial sponsorship of the U.S. Environmental Protection Agency and covers the period from November 1973 to July 1975.

CONTENTS

Page

Foreword iii
Abstract iv
Figures vi
Tables viii
Abbreviations and Symbols xi
Acknowledgements xvi

1. Introduction 1
2. Conclusions 6
3. Recommendations 9
4. Stoichiometric and Kinetic Considerations 12
5. Analytical Methods and Sampling Procedures 34
6. Treatment Facilities, Operating Procedures, and Plant Measurement Procedures 36
7. Raw Sewage Characteristics 46
8. Results and Discussion 51
9. Design Considerations for Activated Sludge Systems Employing Compartmentalized Aeration Tanks 126

References 137
Publications and Patents 140
Appendices 141

FIGURES

Number		Page
1	Coxsackie Experimental Sewage Treatment Plant-Plan of Facilities	2
2	Numbering Scheme for Aeration Cells	4
3	Aeration Patterns in Compartmentalized Reactors	19
4	Two-Step Feeding of Compartmentalized Reactors	21
5	Effluent Ammonia Concentration (S_1) vs. Imposed Nitrifier Growth Rate (μ_{nr})	25
6	Ammonia Profiles in a Plug Flow Reactor Illustrating Table 6	31
7	Surge Tank	37
8	Sideweir Structure	39
9	Section View of Aeration Tank	41
10	Section View of Final Settler	41
11	Chlorine Contact Chamber/Second-Phase Settler Combination	43
12	Raw Sewage Flow Pattern (11/9/73-12/9/73)	49
13	Process Schematic for Flow Sheets I and II	52
14	Aeration Train Dissolved COD Profiles for Flow Sheet I	55
15	Aeration Train Dissolved Orthophosphorus Profiles for Flow Sheet I	56
16	Daily Final Settler Effluent Total COD Profiles for Flow Sheet I	58
17	Daily Final Settler Effluent $NO_2^-+NO_3^-$-N Profiles for Flow Sheet I	59
18	Process Schematic for Flow Sheet III	63
19	Process Schematic for Flow Sheet IV	67
20	Process Schematic for Flow Sheet V	68
21	Process Schematic for Flow Sheet VI	69
22	Monthly Nitrogen Balance Bar Diagrams for Flow Sheets IV, V, and VI	88
23	Flow Sheet I Nitrogen Profiles (11/14/73)	96
24	Flow Sheet I Nitrogen Profiles (11/27/73)	97

FIGURES (continued)

Number Page

25 Flow Sheet I Nitrogen Profiles (12/5/73 98

26 Flow Sheet IV Nitrogen Profiles (9/18/74) 99

27 Flow Sheet IV Nitrogen Profiles (10/10/74) 100

28 Flow Sheet IV Nitrogen Profiles (11/14/74) 101

29 Flow Sheet VI Nitrogen Profiles (6/12/75) 102

30 Flow Sheet I Ammonia and Nitrate Bar Diagram Profiles (11/14/73). 103

31 Flow Sheet I Ammonia and Nitrate Bar Diagram Profiles (11/27/73). 104

32 Flow Sheet I Ammonia and Nitrate Bar Diagram Profiles (12/5/73).. 105

33 Flow Sheet VI Ammonia and Nitrate Bar Diagram Profiles (6/12/75). 106

TABLES

Number		Page
1	Coxsackie Experimental Sewage Treatment Plant - Main Dimensions of Major Treatment Units and Characteristics of Equipment.......	3
2	Characteristic Features of Flow Sheets I Through VI	5
3	Endogenous Nitrate Respiration Stoichiometric Relationships	16
4	Stoichiometric Projection of Nitrogen Removal via Sludge Synthesis and Endogenous Nitrate Respiration	17
5	Substrate Nitrate Respiration and Anoxic Synthesis	19
6	Effect on Basic Kinetic Values of Applying Safety Factor of 2 to Plug Flow Reactor	30
7	Flow-Weighted, Annual-Average Raw Sewage and Surge Tank Effluent Characteristics of Year Ending July 31, 1975 (Flow Sheets IV, V, and VI)	47
8	Flow-Weighted, Monthly-Average Raw Sewage Characteristics for 12 Months Ending July 31, 1975 (Flow Sheets IV, V, and VI)	48
9	Raw Sewage Characteristics for November and December 1973 (Flow Sheet I)	50
10	Average Values of Operating Parameters for Flow Sheet I	53
11	Average Chlorine Contact Chamber Effluent Values for Flow Sheet I	54
12	Operating and Effluent Data for Flow Sheet II	62
13	Operating and Effluent Data for Flow Sheet III	65
14	Air Flow to Plant and DO in Aeration Cells During Flow Sheet IV, V, and VI Operations	70
15	Monthly Hydraulic Surge Tank, Primary Settler, and Chlorine Contact Chamber Data for Flow Sheets IV, V, and VI	71
16	Monthly-Average Aeration Tank and Final Settler Detention Times for Flow Sheets IV, V, and VI	72
17	Monthly-Average Final Settler Overflow and Solids Loading Rates for Flow Sheets IV, V, and VI	73
18	Flow-Weighted, Monthly-Average Surge Tank Effluent Characteristics for Flow Sheets IV, V, and VI	74

TABLES (continued)

Number		Page
19	Monthly-Average Aeration Tank Loading Rates and Solids Retention Times for Flow Sheets IV, V, and VI	75
20	Monthly-Average Mixed Liquor Characteristics for Flow Sheets IV, V, and VI	76
21	Monthly-Average Excess Activated Sludge Characteristics for Flow Sheets IV, V, and VI	77
22	Monthly-Average Solids Production Data for Flow Sheets IV, V, and VI	78
23	Flow-Weighted, Monthly-Average Final Settler Effluent Characteristics for Flow Sheets IV, V, and VI	79
24	Monthly-Average COD and BOD_5 Removal Efficiencies for Flow Sheets IV, V, and VI	80
25	Monthly Treatment System Nitrogen Balances (Excluding Surge Tank and Chlorine Contact Chamber) for Flow Sheets IV, V, and VI	81
26	Monthly Treatment System Fe^{3+} Balances (Excluding Chlorine Contact Chamber) for Flow Sheets IV, V, and VI	82
27	Monthly Treatment System Phosphorus Balances (Excluding Surge Tank and Chlorine Contact Chamber) for Flow Sheets IV, V, and VI	83
28	Effect of Ferric Chloride Addition on Process Phosphorus Removal and Iron Concentration in Return Sludge: Flow Sheet IV	84
29	Flow-Weighted, Monthly-Average Chlorine Contact Chamber Effluent Characteristics for Flow Sheet IV	85
30	Flow-Weighted, Monthly-Average Primary Settler Effluent Characteristics for Flow Sheets V and VI	86
31	Flow-Weighted, Monthly-Average Chlorine Contact Chamber Effluent Characteristics for Flow Sheets V and VI	87
32	Average Contaminant Load on Aeration Train During Flow Sheet IV and VI Operations	93
33	Comparison of Sludge Production and Organic Load Data for Flow Sheets IV and VI	94
34	Stoichiometric Data Abstracted from Nitrogen Profiles for Flow Sheets I, IV, and VI	109
35	Kinetic Data Abstracted from Nitrogen Profiles for Flow Sheets I, IV, and VI	111
36	Monthly Rates of Denitrification Abstracted from Monthly Nitrogen Balances for Flow Sheets IV, V, and VI	113

TABLES (continued)

Number Page

37 Monthly Actual vs. Predicted Volatile Sludge Production for Flow Sheets IV, V, and VI 115

38 Monthly Growth Rates Abstracted from Monthly Nitrogen Balances for Flow Sheets IV, V, and VI 116

39 Monthly-Average Return Sludge Characteristics for Flow Sheets IV, V, and VI 118

40 Effect of Chlorine Contact Chamber/Second Phase Settler Combination on Suspended Solids Removal 119

41 Chlorine Dosages Used During Sludge Chlorination Experiments 122

42 Effect of Sludge Chlorination on Solids Concentration, Filterability, and pH 123

43 Effect of Sludge Chlorination on Slurry Filtrate Quality 124

44 Quality of Sludge Dewatering Bed Underflow During Sludge Chlorination Experiments 125

45 Forms of Nitrogen and Mass of Newly Formed Biomass as Affected by Residence in the Five Zones of the Reactor of the Process Design Example 135

ABBREVIATIONS AND SYMBOLS

ABBREVIATIONS

alk.	alkalinity
A.S.	activated sludge
CC	chlorine contact
CCC	chlorine contact chamber
CM	completely mixed
CRT	cell residence time
diss.	dissolved
DO	dissolved oxygen
EAS	excess activated sludge
ENR	endogenous nitrate respiration
EOR	endogenous oxygen respiration
EPA	U.S. Environmental Protection Agency
EPA PDM	Process Design Manual for Nitrogen Control, published by EPA
F/M	food-to-microorganism ratio
HM	heterotrophic matrix
MCRT	mean cell residence time
ML	mixed liquor
MLSS	mixed liquor suspended solids; mixed liquor suspended solids concentration
MLVSS	mixed liquor volatile suspended solids; mixed liquor volatile suspended solids concentration
N	nitrogen
n.	nitrifier or nitrifiers
N.AP.	not applicable
N.AV.	not available
OUR	oxygen uptake rate
P	phosphorus
PF	plug flow

ABBREVIATIONS AND SYMBOLS (continued)

Q.R.	questionable result
RSSS	return sludge suspended solids; return sludge suspended solids concentration
RSVSS	return sludge volatile suspended solids; return sludge volatile suspended solids concentration
SF	safety factor
SRT	solids residence time ($\equiv$ CRT)
SS	total suspended solids; total suspended solids concentration
SVI	sludge volume index
SWD	side water depth
VSS	volatile suspended solids; volatile suspended solids concentration

SYMBOLS USED IN EQUATIONS
(units given in parentheses)

b	Fractional rate of biomass destruction in a particular endogenous respiration reaction (mg of biomass destroyed per mg of initial biomass).
$b(\text{day}^{-1})$	Microorganism decay factor in endogenous oxygen respiration (mg/mg/day).
b_d	Biomass destruction factor in endogenous nitrate respiration (mg biomass destroyed/mg NO_2^-+NO_3^--N gasified).
C_m(mg/ℓ)	Monthly flow weighted average of any contaminant concentration ($C_m = \Sigma\, C_n Q_n / \Sigma\, Q_n$).
C_n(mg/ℓ)	Contaminant concentration, 24-hr composite, n^{th} day of the month.
E_{ENR}	Stoichiometric efficiency factor in endogenous nitrate respiration (mg N removed from process water/mg NO_2^-+NO_3^--N gasified).
f	Fraction of nitrifier biomass in total biomass of mixed liquor or return sludge.
f_e	Fraction of substrate used for energy in a biological substrate removal process.
Fe(%)	Percent of Fe found in mixed liquor or return sludge solids.
f_s	Fraction of substrate used for synthesis of new biomass in a biological substrate removal process.
$k_d(\text{day}^{-1})$	Specific endogenous nitrate respiration reaction rate (mg NO_2^-+NO_3^--N gasified/mg MLVSS/day).
K_s(mg/ℓ)	Michaelis-Menten constant in nitrification corresponding to prevailing temperature (mg/ℓ NH_4^+-N).

ABBREVIATIONS AND SYMBOLS (continued)

$\dot{n}$(mg/ℓ)	Mass of nitrifiers wasted per liter of reactor space.
$\dot{N}$(mg/ℓ)	Mass of nitrifiers present per liter of reactor space.
$\Delta\dot{N}$(mg/ℓ) or $\Delta\dot{N}_p$(mg/ℓ)	Nitrifier biomass produced per liter of sewage treated.
N_d(mg/ℓ)	Nitrogen gasified in endogenous nitrate respiration per liter of sewage treated.
N_1(mg/ℓ)	Effluent biodegradable nitrogen concentration.
N_0(mg/ℓ)	Influent biodegradable nitrogen concentration.
$\dot{N}_0$(mg/ℓ)	Mass of nitrifiers present per liter of reactor space at time $t = 0$.
$\Delta\dot{N}_w$(mg/ℓ)	Nitrifier biomass wasted per liter of sewage treated.
Q(ℓ/day)	Plant flow.
Q_n(ℓ/day)	Plant flow, n^{th} day of month.
Q_r(ℓ/day)	Return sludge flow.
Q_w(ℓ/day)	Waste sludge flow.
r	Sludge recycle ratio ($r = Q_r/Q$).
S(mg/ℓ)	Substrate concentration.
ΔSC(mg/ℓ)	BOD_5 removed in a process per liter of sewage treated.
SDT(day)	Sewage detention time in entire reactor (SDT = V/Q).
SDT_d(day)	Sewage detention time in denitrification (anoxic) reactor ($SDT_d = V_d/Q$).
SDT_n(day)	Sewage detention time in nitrification (aerobic) reactor ($SDT_n = V_n/Q$).
SDT_p(day)	Sewage detention time in purge cell of aerobic-anoxic reactor ($SDT_p = V_p/Q$).
SF_d	Safety factor, denitrification reactor.
SF_n	Safety factor, nitrification reactor.
S_i(mg/ℓ)	Initial NH_4^+-N concentration after dilution of S_0 by the recycle flow: $S_i = (S_0 + r\,S_1)/(1 + r)$ (In the influent pipe of a reactor or at the head end of a plug flow tank).
S_1(mg/ℓ)	NH_4^+-N concentration in the effluent of a reactor.
S_0(mg/ℓ)	NH_4^+-N concentration in the influent wastewater.
SS_{SV30}(mg/ℓ)	Suspended solids concentration in the space occupied by settled sludge at the end of a 30-min settled volume test.
SV30(mℓ)	Volume occupied by sludge at the end of a 30-min settling test for a 1-liter test volume of mixed liquor.

ABBREVIATIONS AND SYMBOLS (continued)

SV60(mℓ)	Volume occupied by sludge at the end of a 60-min settling test for a 1-liter test volume of mixed liquor.
SVI(mℓ/g)	Sludge volume index.
t(day)	Time.
t_m(day)	Flowthrough time in plug flow nitrification reactor necessary to achieve prescribed effluent NH_4^+-N concentration.
t_n(day)	Flowthrough time in plug flow nitrification reactor $[t_n = SDT_n/(1 + r)]$.
T(C)	Temperature.
v	Fractional volatility of mixed liquor solids (v = VSS/SS); final fractional volatility of mixed liquor solids at the end of a particular endogenous respiration period.
V(ℓ)	Volume of reactor.
V_d(ℓ)	Volume of denitrification (anoxic) reactor.
V_n(ℓ)	Volume of nitrification (aerobic) reactor.
v_0	Fractional volatility of mixed liquor or return sludge solids at the beginning of endogenous respiration.
Vol. (%)	Percentage of volatility of mixed liquor solids or of return sludge solids (Vol. = 100 v).
V_p(ℓ)	Volume of purge compartment of aerobic-anoxic reactor.
$w_n(\text{day}^{-1})$	Specific nitrifier biomass wasting rate (mg/mg/day).
X(mg/ℓ)	Mixed liquor volatile suspended solids concentration (biomass concentration).
ΔX(mg/ℓ)	Net biomass production (mg VSS synthesized less mg VSS destroyed in EOR and ENR) per liter of sewage treated.
ΔX_w(mg/ℓ)	Biomass wasted (mg VSS) per liter of sewage treated.
Y	Yield factor, BOD_5 removal (mg VSS produced/mg BOD_5 removed at zero cell residence time).
Y'	Adjusted aerobic yield factor, BOD_5 removal (mg VSS produced/mg BOD_5 removed at zero cell residence time less aerobic decay: $Y' = Y - b \times SDT_n/\Delta SC$).
Y_n	Yield factor, nitrification (mg nitrifier biomass produced/mg NH_4^+-N oxidized).
Θ	Theta, base of exponential factor used as temperature coefficient in the Arrhenius-Phelps-Streeter equation that describes the effect of temperature on rate of biological reactions.
$\mu_{hr}(\text{day}^{-1})$	Specific growth rate of the heterotrophic matrix prevailing in the system ($\mu_{hr} = \Delta X/X\ SDT$).

ABBREVIATIONS AND SYMBOLS (continued)

$\mu_{max}(day^{-1})$	Maximum specific nitrifier growth rate, within the context of the Monod function, corresponding to prevailing temperature, DO, and pH.
$\mu_n(day^{-1})$	Specific instantaneous nitrifier growth rate corresponding to prevailing NH_4^+-N concentration and environmental conditions.
$\mu_{na}(day^{-1})$	Average specific nitrifier growth rate computed for one reaction cycle of a plug flow reactor.
$\mu'_{na}(day^{-1})$	Average specific nitrifier growth rate in a plug flow reactor computed for one reaction cycle, neglecting effect of sludge recirculation.
$\mu_{nr}(day^{-1})$	Specific nitrifier growth rate to be imposed on system in accordance with effluent requirements and applicable process kinetics.
$\mu_{nd}(day^{-1})$	Specific nitrifier growth rate selected by designer for imposition on the system ($\mu_{nd} = \mu_{nr}/SF_n$).
Σ	Summation over the days of 1 mo.

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SECTION 1

INTRODUCTION

This project was initiated in the summer of 1966 when the New York State Departments of Correction and Health agreed to undertake a program of research and development in the area of sewage treatment at one of its correctional facilities. With a view to the impending program of sewage treatment plant construction in New York State, it was decided to build an experimental sewage treatment plant and a research laboratory on the grounds of the Coxsackie Correctional Facility at West Coxsackie, New York. The plant was intended to treat the wastewater of that penal facility and to be devoted to research in the area of activated sludge. It was felt that the activated sludge process would be the principal process underlying the plants to be constructed in the State. Assistance in conducting this research program was sought from the Federal Water Pollution Control Administration, and a grant application was finally approved by that agency in July 1968.

The Coxsackie Correctional Facility is an institution for young male delinquents aged 16 to 21. The average age of the inmates is 18. Design inmate population is 750. During the period covered by the full analytical data of this report, the inmate population was near capacity. In addition to the inmates, approximately 350 prison personnel are in daytime or nighttime residence at the facility. A 306-ha (750-ac) farming operation is part of the correctional facility. Farm products are milk, vegetables, apples, and beef. The institution is located on U.S. Route 9W, 39 km (24 mi) south of Albany. The effluent of the sewage plant is discharged to the Coxsackie Creek, a short tributary of the Hudson River. The Coxsackie Creek is classified as an intermittent stream. The New York State effluent requirements for sewage treatment plant discharging to intermittent streams are as follows:

BOD_5	max. 5 mg/ℓ
NH_4^+-N	max. 2 mg/ℓ
DO	min. 7.5 mg/ℓ

The facilities which comprised the research installation are shown in Figure 1. The main dimensions of the major treatment units employed on this project are summarized in Table 1.

The experimental sewage treatment plant was placed in operation in July 1972. The first year of operation did not furnish any data because of frequent operational catastrophies resulting from the breakdown of improperly installed

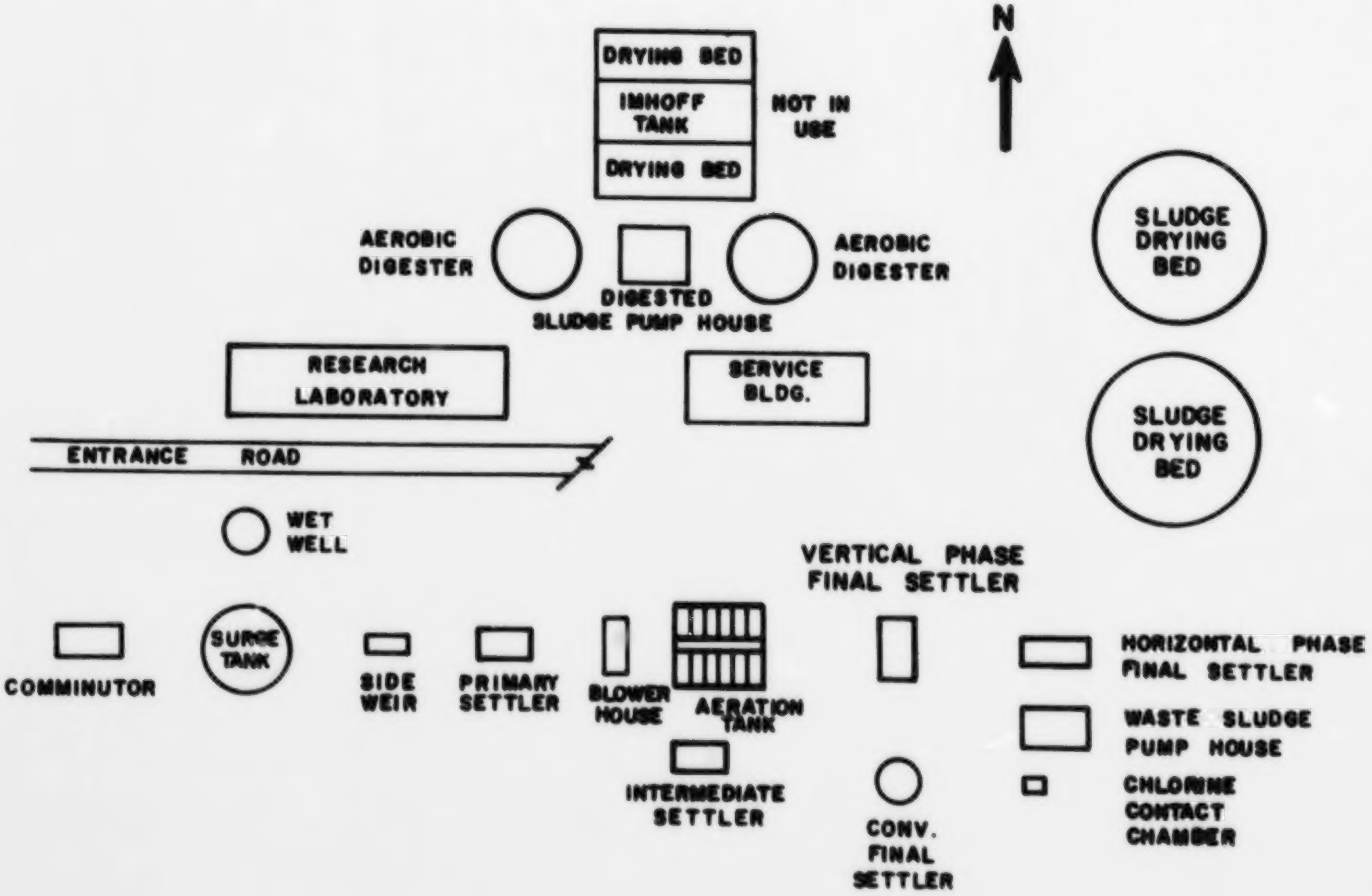
N
DRYING BED
IMHOFF TANK
DRYING BED
NOT IN USE
AEROBIC DIGESTER
DIGESTED SLUDGE PUMP HOUSE
AEROBIC DIGESTER
SLUDGE DRYING BED
RESEARCH LABORATORY
SERVICE BLDG.
SLUDGE DRYING BED
ENTRANCE ROAD
WET WELL
VERTICAL PHASE FINAL SETTLER
COMMINUTOR
SURGE TANK
SIDE WEIR
PRIMARY SETTLER
BLOWER HOUSE
AERATION TANK
INTERMEDIATE SETTLER
CONV. FINAL SETTLER
HORIZONTAL PHASE FINAL SETTLER
WASTE SLUDGE PUMP HOUSE
CHLORINE CONTACT CHAMBER

Figure 1. Coxsackie experimental sewage treatment plant - plan of facilities.

TABLE 1. COXSACKIE EXPERIMENTAL SEWAGE TREATMENT PLANT - MAIN DIMENSIONS OF MAJOR TREATMENT UNITS AND CHARACTERISTICS OF EQUIPMENT

AVERAGE DAILY FLOW:

570 m^3 (0.15 mgd)

SCREENS:

Two manually raked screens in series; 25 and 13 mm spacing of bars

GRIT REMOVAL:

Some grit removal occurs in surge tank and flow control structure (Side weir structure)

SURGE TANK:

9.76 m diameter; approximately 215 m^3 working capacity; equipped with floating aerator with direct-coupled, propeller-type impeller, 7.45 kw/1150 rpm; outpumping by submersible fixed-speed, torque-flow pumps to flow control structure

CONTROL OF FLOW TO TREATMENT UNITS:

Manual, by varying opening of submerged orifice in constant head device

PRIMARY SETTLER (Optional Use):

1.83 m wide; 7.32 m long; 2.44 m sidewater depth; 32.7 m^3 volume; 13.4 m^2 area; 1.83 m weir

AERATION TANK:

12 compartments, each 1.53 x 3.05 m in plan; sidewater depth 2.31 or 2.61 m, adjustable; volume under aeration 129 or 146 m^3; INKA aeration grids mounted near floor, 3.2 mm diam. airholes; one double compartment equipped with 0.25 kw/350 rpm mixer; compartments in series

AIR BLOWER:

Centrifugal type; 3.2 to 14.16 m^3/min; 15 kw

FINAL SETTLING TANK:

Peripheral entry; 4.88 m diameter; 3.05 m sidewater depth; 57 m^3 volume; 18.70 m^2 area, 12.81 m weir

SLUDGE RETURN PUMPS:

Plunger Type

CHLORINE CONTACT CHAMBER:

15.9 m^3 volume

equipment. Also, operations were somewhat handicapped by a sludge disposal problem.

In the fall of 1973, Flow Sheet No. I was implemented as the first of a series of six flow sheets evaluated, all characterized by the use of anoxic zones in a plug flow (PF) aeration tank and by the use of ferric chloride to enhance phosphorus removal and flocculation. The main emphasis of this research was on nitrogen removal without use of an external organic carbon source. Table 2 contains a description of the six flow sheets evaluated, and Figure 2 depicts aeration cell numbering.

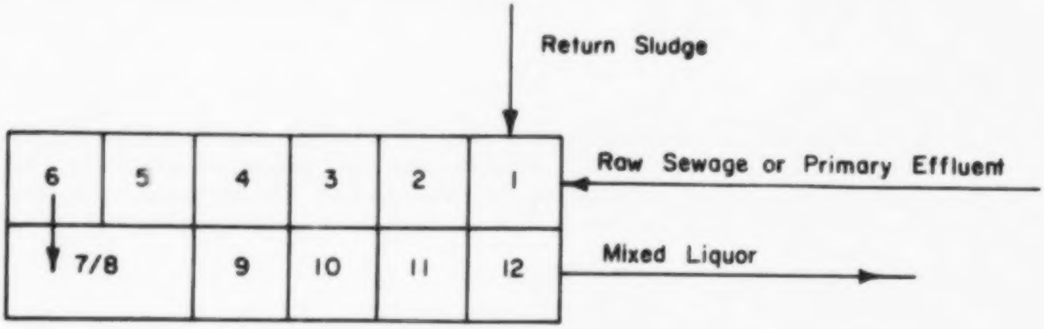

Figure 2. Numbering scheme for aeration cells.

A PF regime was achieved in the aeration tank by providing 11 compartments or cells arranged in series. Raw sewage or primary effluent was introduced into Cell 1 along with the return sludge. Ten of the 11 compartments were of equal size; Cell 7/8 was double the size of the other compartments.

TABLE 2. CHARACTERISTIC FEATURES OF FLOW SHEETS I THROUGH VI

	Period	Surge Tank	Primary Settler	Scheme of Aeration
I Fig. 13	Nov. 1973 to March 1974 inclusive	Anoxic, sludge scraper used	In use	Return sludge & primary effluent introduced into Cell 1 - Cells 1, 7/8, 9, 10, 11 anoxic - $FeCl_3$ dosed to primary effluent (1)
II Fig. 13	April 26 to May 6, 1974	Aerobic, floating aerator on full time	In use	Same as Flow Sheet I - Cells 1, 7/8, 9, 10, 11 anoxic (1)
III Fig. 18	May 7 to June 21, 1974	Used as part of the activated sludge system	By-passed	Return sludge & raw sewage introduced into surge tank - Cells 5, 6, 7/8, 9, 10 anoxic - $FeCl_3$ dosed to ML leaving surge tank (1)
IV Fig. 20	June 22, 1974 to March 31, 1975	Aerobic, floating aerator on 2 min, off 18 min	By-passed	Return sludge and raw sewage introduced into Cell 1 - Cells 5, 6, 7/8, 9, 10 anoxic - $FeCl_3$ dosed to either surge tank effluent and/or ML leaving aeration tank (2)
V Fig. 21	April 1975	Same as Flow Sheet IV	In use	Return sludge & primary effluent introduced into Cell 1, otherwise same as Flow Sheet IV (2)
VI Fig. 22	May to July 1975 inclusive	Same as Flow Sheet IV	In use, in conjunction with chemical addition	Return sludge & primary effluent introduced into Cell 1 - same aeration pattern as Flow Sheet IV but also alternating aerobic and anoxic zones - $FeCl_3$ addition to primary influent & ML leaving aeration system (2)

(1) Aerator Volume used: 129 m^3
(2) Aerator Volume used: 146 m^3

SECTION 2

CONCLUSIONS

Three forms of nitrate respiration may be postulated to occur in single-sludge activated sludge (A.S.) systems designed for nitrogen removal. In all three forms, the reaction is mediated by the heterotrophic microbial biomass that makes up the bulk of the mixed liquor suspended solids (MLSS) of the system, to wit:

a) Substrate nitrate respiration: the interaction, under anoxic conditions, of organic wastewater carbon compounds with nitrates, the carbon compounds serving as the hydrogen donor. Organic wastewater carbon compounds are present in the liquid phase of the process water.

b) Endogenous nitrate respiration (ENR): the interaction, under anoxic conditions, of the heterotrophic biomass itself with the nitrates, the biomass serving as the hydrogen donor. Biodegradable carbon compounds are absent from the liquid phase of the process water.

c) Adsorbed carbon nitrate respiration: the interaction, under anoxic conditions, of organic carbon compounds adsorbed by the biomass with nitrates, the adsorbed matter serving as the hydrogen donor. Biodegradable carbon compounds are absent from the liquid phase of the process water.

An in-line surge tank, operated as part of an A.S. system and subjected to pulsating aeration, can be used to stimulate substrate nitrate respiration to occur directly in the surge tank. Pulsating aeration may be provided by a floating, direct-drive surface aerator, operated on a 10 min "on", 10 min "off" cycle.

ENR can be effected in an A.S. system by providing a compartmentalized reactor and operating the compartments to produce an aerobic-anoxic-aerobic pattern of aeration for a PF hydraulic regime. The aerobic period at the effluent end of the reactor is provided to nitrify any ammonia released by the biomass in the preceding anoxic compartments and to purge the gaseous nitrogen that clings to the A.S. floc.

Adsorbed carbon nitrate respiration seems to occur in conjunction with ENR if the preceding aerobic period is kept as short as possible. The following additional conclusions can be drawn with respect to ENR:

1) ENR may be used instead of endogenous oxygen respiration (EOR) for biomass destruction and for conditioning the A.S. floc for liquid-solids

separation in the final settler. Analysis of sludge production data under this project points to a biomass destruction rate of 2.2 mg/mg N removed.

2) ENR is accompanied by generation of alkalinity, in part compensating for destruction of alkalinity during nitrification. Five alkalinity and nitrogen profiles through the compartmentalized reactor developed during the project indicated alkalinity generation of 2.6 to 3.9 mg/mg NO_3^--N gasified, alkalinity being expressed as $CaCO_3$.

3) ENR is accompanied by ammonia gas release. The extent of that release varies considerably with the treatment situation and is generally far below the extent predicted by stoichiometric theory. Based on an assumed biomass composition of $C_5H_7O_2N$, the stoichiometric release of ammonia in ENR is 0.25 mg/mg NO_3^--N gasified. Five nitrogen profiles through the compartmentalized reactor, developed for three different treatment schemes, indicated ammonia release varying between 0.03 and 0.17 mg/mg NO_3^--N gasified. It was concluded that adsorbed carbon nitrate respiration contributed to the difference between actual and theoretical NH_4^+-N release and that the assumed biomass composition formula - $C_5H_7O_2N$ - might not be correct.

4) For a particular pattern of aeration, the specific nitrate reduction rate (mg NO_3^--N gasified to N_2/mg MLVSS/day) is:

   - independent of cell residence time (CRT) within the 4- to 9-day range used on the project,
   - independent of nitrate nitrogen concentration down to 1 mg/ℓ NO_2^- + NO_3^--N, and
   - independent of the MLSS concentration within the range used in this project, i.e., 2500-4000 mg/ℓ.

5) Within the ranges indicated in the preceding paragraph, ENR may be regarded as a zero order reaction with respect to NO_2^- + NO_3^--N and MLVSS.

6) Incorporation of ENR into an A.S. process does not necessarily affect nitrification unfavorably.

With respect to nitrification in a single-sludge A.S. system in a compartmentalized reactor, the following conclusions are based on observations made during the project:

1) At the very beginning of the aerobic reaction period, a short interval of inhibition of nitrate generation occurs. However, ammonia removal is not impaired. This phenomenon cannot be observed by examining the ammonia profile; it is visible in the nitrate profile. The phenomenon does not seem to be caused by a deficiency of dissolved oxygen (DO) in the process water. During this period of inhibition, NH_4^+-N is taken up by the heterotrophic biota.

2) Alkalinity destruction during nitrification is very near the theoretical (stoichiometric) value of 7.14 mg/mg NH_4^+-N nitrified.

3) Except for the short inhibition period referred to above, the nitrate profile has the shape of a Monod curve, as generally maintained in the literature (1).

4) Providing a larger CRT for the heterotrophic matrix will increase the amount of nitrifiers and the rate of nitrification, as predicted by kinetic theory.

5) The monthly nitrifier growth rates computed on the basis of the growth rates of the heterotrophic matrix imposed on the A.S. system appear to be considerably higher than the growth rates that should have occurred in accordance with kinetic theory and kinetic coefficients found in the literature.

The use of a compartmentalized aeration tank allows the allocation of treatment time to aerobic and anoxic periods in accordance with seasonal requirements.

Excessive preaeration of untreated sewage in a surge tank results in a turbid (milky) effluent from the downstream A.S. system.

Chlorination of sludge (both excess activated and primary) to a pH of 2.5 in a pipe reactor of 4-min detention time will effectively improve drainage properties of the sludge to an extent comparable to that obtainable in commercial sludge chlorination reactors.

Under aerobic conditions, heterotrophic bacteria prefer organically bound nitrogen to ammonia as a source of nitrogen.

SECTION 3

RECOMMENDATIONS

When designing sewage treatment plants for nitrogen removal, consideration should be given to single-sludge A.S. systems utilizing ENR as a means of denitrification and biomass destruction. The following detailed recommendations apply to the design of such plants:

1) Compartmentalized aeration tanks should be utilized with provisions for both the aerobic as well as anoxic mode of operation in most compartments. Compartmentalization will achieve an approximation of PF, which from a kinetic standpoint is a more efficient hydraulic regime than a completely mixed (CM) regime. In contrast to theoretical PF, compartmentalization will provide a measure of damping of shock loads depending upon the size of the first compartment. Compartmentalization also will provide flexibility of the aeration pattern allowing adaptation to fluctuating loads and reaction rates.

2) The last compartment in a compartmentalized reactor should be an aerobic compartment for the stripping of nitrogen gas bubbles clinging to the A.S. floc and for nitrifying the ammonia released in the preceding anoxic compartments.

3) If maximum utilization of wastewater carbon compounds for nitrogen removal is intended, the plant should be designed without primary settling. This will make more biomass available for the $NO_2^- + NO_3^-$-N destruction that occurs in ENR.

4) To consistently achieve an effluent BOD_5 concentration of 10 mg/l or less, an overall mean cell residence time (MCRT) of at least 10 days should be provided for. At shorter CRTs, it is difficult to achieve the required good flocculation of the A.S. floc.

Experience from this project points to the need for additional kinetic research in the area of nitrification. The average monthly nitrifier growth rates observed differed from those predicted for the prevailing treatment conditions on the basis of recommended kinetic theory and coefficients. Also, the nitrate profiles through a PF reactor were not found to fully follow the Monod function, stipulated frequently as the relationship governing nitrification. These discrepancies are understandable when considering that the present kinetic theory of nitrification is largely built on the behavior of pure cultures maintained in river (Thames) water. The bulk of the recommended additional research should be undertaken with mixed cultures (derived from municipal sewage treatment plants) using CM reactors. By keeping the wasting rates for the reactors constant and maintaining uniform environmental conditions and ammonia nitrogen loads, it should be possible to obtain realistic growth rate data as a function of prevailing substrate concentration and environmental conditions which would be applicable to sewage treatment plants. Kinetic research on CM reactors should be supplemented by research in fill-and-draw reactors also using mixed cultures. This research should lead to an improved functional equation relating the effluent ammonia concentration of PF reactors to the imposed growth rate of the nitrifiers. That equation would include influent ammonia nitrogen (N) concentration, influent BOD_5 concentration, and the ratio of sludge recirculation to sewage flow as variable operational parameters.

There is a dearth of data supporting yield coefficients for aerobic respiration/synthesis and nitrate respiration/synthesis. Yet these yield coefficients are vitally important for the sizing of single-sludge A.S. plants. Research should be undertaken with fill-and-draw reactors to remedy this deficiency. In the aerobic process, the onset of biomass destruction via EOR may be pinpointed by observing the substrate (TOC) and ammonia profiles, provided nitrification is suppressed. Operating the process in this manner should not only furnish yield data but also data regarding initial sludge volatility, another kinetic coefficient for sizing and operating single-sludge A.S. plants. In the anoxic process, disappearance of nitrates will pinpoint the end of substrate nitrate respiration.

Further research is recommended with respect to the biomass destruction (decay) coefficient prevailing with PF regimes under both aerobic and anoxic conditions at various temperatures. This coefficient is needed for sizing single-sludge A.S. plants designed for N removal. This research should be conducted in fill-and-draw reactors which simulate PF systems. Sludge filterability should be determined at the end of the reaction; this is an important factor for judging cost effectiveness of a process. Results of this research would also furnish the following data, all important in sizing and operating the reactors: percent of biomass destruction as a function of percent sludge volatility, practical limits of biomass destruction, biomass destruction as a function of NO_3^--N gasification, and ammonia and alkalinity release in various treatment schemes.

Promising results were achieved using a chlorine contact chamber (CCC) as a second-phase final settler. A multiple-tray cross flow configuration was employed. Further research and development work to perfect this concept is recommended. The essence of this work should be the development of suitable corrugated profiles for the trays of the settler. Flat sheets were used on this project.

Experience gained on this project points to greater SS and BOD removal capacities for A.S. plants employing nitrate respiration as compared with single-sludge A.S. plants of the fully aerobic type. However, sufficient time was not available to produce quantitative data reflecting the effect of nitrate respiration on removal efficiencies. Research in parallel reactors to produce such data should be undertaken. Incorporation of a certain measure of nitrate respiration in all A.S. processes seems indicated. The extent of this measure should also be a subject of the suggested parallel-reactor research.

Further research is also recommended concerning the utilization of a surge tank as part of an overall A.S. system operated for nitrogen removal. The surge tank should be operated in a pulsating aeration mode, thus stimulating substrate nitrate respiration to take place directly in the surge tank. The surge tank, in turn, should be followed by a compartmentalized reactor employing ENR. Excellent overall system removals were achieved with this arrangement during the project; however, the investigation could not be completed because of sludge bulking.

The BOD and SS removal capabilities of an A.S. process is directly related to the flocculating characteristics of its MLSS. For plant operating purposes, as well as for comparative purposes, it would be desirable to have a simple flocculating test available. Such a test should be developed and standardized. It could consist simply of determining the SS concentration in the supernatant of a suitably sized laboratory settler after 30 min of settling.

In the course of this study, it became clear that two basic patterns of aeration are available for the operation of single-sludge A.S. systems in compartmentalized reactors: the "block" pattern and the "alternating zones" pattern. A comparative evaluation of the two approaches under conditions of changing loads and reaction rates is recommended. The "alternating zones" approach may require less operator adjustment to adapt the process to changing loads and reaction rates; however, project results related to this subject are inconclusive.

SECTION 4

STOICHIOMETRIC AND KINETIC CONSIDERATIONS

In this section, certain aspects of single-sludge A.S. systems using nitrate respiration are discussed in order to present a working theory for the processes investigated under this project. The discussion is based on data found in the literature, particularly in the Process Design Manual for Nitrogen Control (1) published by the U.S. Environmental Protection Agency (EPA PDM). The material covered in this section does not coincide with the scope of the related sections of the EPA PDM, and the approach taken here differs in many respects from the approach of that manual.

Section 4 represents an attempt to postulate models of certain complex biological processes. As in other attempts of this nature, a good measure of idealization and simplification had to be used.

The term "substrate" as used here, and throughout this project report, refers to matter that is dissolved or suspended in the liquid phase of the process water. "Substrate" does not refer to matter already adsorbed to the surface of the biomass.

The term "anoxic" refers to the DO concentration of the liquid phase of the process water. Anoxic conditions prevail if the DO of the process water is approximately 0.5 mg/ℓ or less. Nitrates may or may not be present under anoxic conditions.

NITRATE RESPIRATION IN SINGLE-SLUDGE ACTIVATED SLUDGE SYSTEMS

Four routes of nitrogen removal seem to be available in single-sludge A.S. sludge systems. These routes are:

a) Sludge synthesis: this may occur either under aerobic or anoxic conditions. The stoichiometric equation describing sludge synthesis is the same for both conditions. In the course of the synthesis reaction, ammoniacal or organically-bound nitrogen is removed from the substrate by the biomass and incorporated into new biomass. The energy deficiency of the synthesis reaction is covered by the energy surplus of aerobic or nitrate (substrate) respiration.

b) Substrate nitrate respiration: the interaction of nitrates with organic wastewater carbon compounds mediated by the biomass. This reaction always accompanies sludge synthesis under anoxic conditions. Nitrates are utilized as the hydrogen acceptor, organic wastewater carbon compounds as the hydrogen donor. This is an energy reaction. Nitrogen is released to the atmosphere. During this reaction, biodegradable carbon compounds are present in the liquid phase of the process water.

c) Endogenous nitrate respiration: the interaction of nitrates with the sludge biomass itself under anoxic conditions. Nitrates are utilized as the hydrogen acceptor; the biomass itself serves as the hydrogen donor. This too is an energy reaction. Nitrogen is released to the atmosphere and ammonia is released to the process water. During ENR, biodegradable carbon compounds are absent from the liquid phase of the process water. Adsorbed carbon nitrate respiration often accompanies ENR. Because the chemistry of the biomass and of adsorbed carbon compounds is not well known, it is often impossible to differentiate quantitatively between ENR and adsorbed carbon nitrate respiration. In later sections of this report, the term ENR is, therefore, used to cover both denitrification based on biomass destruction and denitrification based on the utilization of adsorbed carbon compounds as the hydrogen donor.

d) Adsorbed carbon nitrate respiration: the interaction, under anoxic conditions, of nitrates with organic carbon compounds adsorbed by the biomass and occurring while the liquid phase of the process water is already devoid of biodegradable carbon compounds. The adsorbed carbon compounds are utilized as the hydrogen donor by the biomass; nitrates serve as the hydrogen acceptor. This too is an energy reaction. Nitrogen is released to the atmosphere. The reaction occurs in conjunction with ENR and is accompanied by a measure of sludge synthesis. The stoichiometry of this reaction is probably similar to that of substrate nitrate respiration, but not enough data are available to allow the postulation of stoichiometric equations. For reactor design purposes, the kinetic equations describing ENR can be easily adapted to reflect the occurrence of adsorbed carbon nitrate respiration. This is shown in following sections. The stoichiometry and kinetics of adsorbed carbon nitrate respiration are not discussed further in this report.

Stoichiometric and Metabolic Assumptions

This stoichiometric discussion rests on the following basic assumptions:

1) The composition of the biomass produced in the A.S. process is $C_5H_7O_2N$ (2).

2) The composition of the carbon source in domestic sewage is $C_{10}H_{19}O_3N$ (3).

3) The energy/synthesis ratio ($f_e:f_s$) for aerobic biological removal of the carbon source is 0.54:1.

4) The energy/synthesis ratio ($f_e:f_s$) for anoxic biological removal of the carbon source is 1.86:1.

5) The BOD_5/COD ratio of the carbon source is 0.45:1.

The $f_e:f_s$ ratio is the ratio of substrate utilized for energy to substrate utilized for synthesis in a reaction of zero CRT, i.e., in a reaction in which no endogenous respiration occurs. The nomenclature $f_e:f_s$ is due to McCarty (3).

The carbon source does not include free ammonia in the domestic sewage. The energy/synthesis ratio for the aerobic process is a rounded version of the ratio stipulated by Porges et al. (2); the f_e:f_s ratio for the anoxic process was estimated on the basis of data reported by Barnard (4).

It is readily conceded that the experimental basis of the assumptions used leaves much to be desired. This applies in particular to the assumptions made regarding biomass composition and the f_e:f_s ratio for the anoxic process. Yet, it was felt that available data are good enough to serve as a basis for a working theory.

From the assumptions listed, several conclusions may be readily drawn, to wit:

1) The yield factor Y based on BOD_5 (mg MLVSS produced/mg BOD_5 removed) in aerobic substrate utilization is 1.02; Y is o.46 in terms of COD.

2) During anoxic substrate utilization, Y is 0.55 in terms of BOD_5, 0.25 in terms of COD.

3) One mole of the carbon source in domestic sewage ($C_{10}H_{19}O_3N$) has a COD of 400 g, a BOD_5 of 180 g, and a TOC of 120 g and weighs 201 g.

4) One mole of the biomass (MLVSS) produced has a COD of 160 g, weighs 113 g, and contains 12.4 percent nitrogen.

5) The stoichiometric equation for sludge synthesis under aerobic as well as anoxic conditions is:

$$C_{10}H_{19}O_3N + 1.5\ NH_3 + 2.5\ CO_2 \longrightarrow 2.5\ C_5H_7O_2N + 3\ H_2O \qquad (1)$$

It should be noted that NH_4^+-N is assumed to be the source of the additional nitrogen needed, even under anoxic conditions.

6) The stoichiometric equation for aerobic (oxygen) respiration in the presence of substrate is:

$$C_{10}H_{19}O_3N + 12.5\ O_2 \longrightarrow 10\ CO_2 + 8\ H_2O + NH_3 \qquad (2)$$

Usually this reaction is simply referred to as "respiration."

7) The stoichiometric equation for substrate nitrate respiration is:

$$C_{10}H_{19}O_3N + 10\ NaNO_3 \longrightarrow 10\ CO_2 + 3\ H_2O + NH_3 + 10\ NaOH + 5\ N_2 \qquad (3)$$

8) The stoichiometric equation for aerobic removal (synthesis + respiration) of the carbon source reads:

$$C_{10}H_{19}O_3N + 4.375\ O_2 + 0.625\ NH_3 \longrightarrow 1.875\ CO_2 + 4.75\ H_2O + 1.625\ C_5H_7O_2N \qquad (4)$$

9) The stoichiometric equation for anoxic removal (synthesis + respiration) of the carbon source reads:

$$C_{10}H_{19}O_3N + 6.5\ NaNO_3 \rightarrow 0.125\ NH_3 + 5.625\ CO_2 + 0.875\ C_5H_7O_2N$$
$$+ 6.5\ NaOH + 3.25\ N_2 + 3\ H_2O \qquad (5)$$

In Equations 3 and 5, $NaNO_3$ stands for all the nitrates present.

Construction of Equations 1 to 5 is based on the half-reaction equations originally proposed by McCarty (3). The heuristic strategy of looking on all biological substrate removal processes as composed of a respiration (energy) reaction and a synthesis (sludge production) reaction was introduced into sanitary engineering by Porges et al. (2).

Additional details on the subject of this subsection may be found in Reference 5.

Endogenous Nitrate Respiration (ENR)

ENR was introduced to sewage treatment in 1964 by Wuhrmann of Zurich (6) who inserted an anoxic reactor between the aeration tank and final settler of a conventional A.S. system (see Figure 5-15 of the EPA PDM). The anoxic treatment tank was equipped with mixing devices to keep the biomass in suspension.

Equations 6 and 7 describe endogenous oxygen respiration (EOR) and ENR, respectively. Equation 6 has been stipulated by Porges et al. (2) and Equation 7 by McKinney (7) and Christensen et al. (8). From Equations 6 and 7, Equations 6a and 7a were derived by considering the reactions between CO_2, NH_3, and NaOH that may be predicted to occur in the process water (for further details, see Reference 5).

$$C_5H_7O_2N + 5\ O_2 \rightarrow 5\ CO_2 + NH_3 + 2\ H_2O \qquad (6)$$

$$C_5H_7O_2N + 5\ O_2 \rightarrow 4\ CO_2 + NH_4HCO_3 + H_2O \qquad (6a)$$

$$C_5H_7O_2N + 4\ NaNO_3 \rightarrow 5\ CO_2 + NH_3 + 2\ N_2 + 4\ NaOH \qquad (7)$$

$$C_5H_7O_2N + 4\ NaNO_3 + H_2O \rightarrow 4\ NaHCO_3 + NH_4HCO_3 + 2\ N_2 \qquad (7a)$$

The same amount of ammonia nitrogen is released in both the EOR and ENR reactions. In the aerobic reaction, the ammonia nitrogen is immediately nitrified. This is not the case in the anoxic reaction. Extra aerobic reactor space must be provided if it is desired to oxidize the ammonia nitrogen formed during ENR. This was not done by Wuhrmann.

The release of ammonia nitrogen by the biomass gives rise to a slight momentary increase in alkalinity. In the anoxic reaction, additional alkalinity is released due to the reduction of nitrates (3.57 mg alkalinity as $CaCO_3$/mg NO_3^--N gasified). This is one-half of the amount used up in the nitrification of 1 mg of ammonia nitrogen. Once the nitrates are exhausted, the plant operator will be confronted with unneeded reactor space if the anoxic zone is too large.

From Equation 7a, the values of Table 3 have been abstracted.

TABLE 3. ENDOGENOUS NITRATE RESPIRATION STOICHIOMETRIC RELATIONSHIPS

1 mg of NO_3^--N gasified	2 mg of biomass destroyed
1 mg of NO_3^--N gasified	0.25 mg of NH_4^+-N released to the liquid phase of the process water due to biomass destruction
1 mg of NO_3^--N gasified	1.07 mg of biomass carbon destroyed
1 mg of NO_3^--N gasified	4.46 mg of total alkalinity produced: 3.57 mg due to reduction of NO_3^--N, 0.89 mg due to biomass destruction
1 mg of NO_3^--N gasified	0.75 mg of N removed from the liquid phase of the process water: 1 mg of NO_3^--N is gasified, but 0.25 mg of NH_4^+-N is added
1 mg of N removed*	2.69 mg of biomass destroyed
1 mg of N removed*	1.43 mg of biomass carbon destroyed
1 mg of N removed*	1.33 mg of NO_3^--N gasified

* From the liquid phase of the process water.

Of particular importance with ENR are two stoichiometric relationships:

a) For 4 mg of NO_3^--N gasified, 1 mg of NH_4^+-N is released back to the process water.

b) Two mg of biomass are destroyed for each mg of NO_3^--N reduced; 2.69 mg of biomass are destroyed for each mg of N removed.

Table 3 concerns a single-stage reaction. All N removed in this reaction from the mixed liquor (ML) is removed by gasification of NO_3^--N. But the amount of NO_3^--N gasified is larger than the net amount of N removed from the liquid phase of the process water. The difference is the ammoniacal N released by the biomass that was destroyed during ENR.

If it is desired to remove the residual ammonia nitrogen generated during ENR, a cascade approach would have to be followed if the basic scheme suggested by Wuhrmann is used. The NH_4^+-N released in the first ENR zone would be nitrified in a following aerobic zone. The nitrate nitrogen generated there would be reduced in a second ENR zone. This would result in a second NH_4^+-N residual, amounting to approximately one-quarter of the first residual. Repeated application of the cascade approach would result in an overall process in which mg of N removed from the liquid phase of the process water would equal mg of NO_3^--N gasified.

Nitrogen Removal by ENR and Aerobic Sludge Synthesis

The effect of increasing rates of biomass destruction via ENR is shown in Table 4. It is assumed that 100 mg of BOD_5 are removed and that the initial sludge volatility is 80 percent. The first column represents a theoretical process with no biomass destruction. Nitrogen removal is 12 mg, due only to sludge synthesis. The 80 percent volatility stays unchanged. In the second column, 42 percent biomass destruction occurs, resulting in a final sludge volatility of 70 percent. A part of this biomass destruction is assumed to be due to EOR, with no benefit for nitrogen removal. Ten percent of the initial biomass of 100 mg, or 10 mg, is assumed to have been lost in this way. The nitrogen removal in this process is 19 mg. If EOR had been used for biomass destruction, the removal would have been 7 mg. The third column represents a process with 63 percent biomass destruction. Ten percent of the original biomass of 100 mg is lost again in EOR. The final sludge volatility is 60 percent; nitrogen removal is 24 mg. If EOR had been used throughout, nitrogen removal would have been 4 mg.

In Table 4, it was assumed that 2.69 mg of biomass were destroyed for each mg of N removed.

TABLE 4. STOICHIOMETRIC PROJECTION OF NITROGEN REMOVAL VIA SLUDGE SYNTHESIS AND ENDOGENOUS NITRATE RESPIRATION *

Final sludge volatility	(%)	80	70	60
Biomass destruction	(%)	-	42	63
Initial biomass	(mg)	100	100	100
Biomass used in ENR	(mg)	-	32	53
Biomass lost in EOR	(mg)	-	10	10
Biomass final	(mg)	100	58	37
N removed in sludge	(mg)	12	7	4
N removed via ENR	(mg)	-	12	20
N removed total	(mg)	12	19	24

* Based on 100 mg of BOD_5 removed and 80 percent initial sludge volatility.

The assumed 80 percent value for initial volatility is probably conservative. The actual value may be somewhat higher; 82 percent volatility, for instance, was reported for the Newtown Creek return sludge at a CRT of 3 days with pure oxygen operation (9). On the other hand, initial volatility might be much lower than 80 percent due to metal salt addition for phosphorus removal. The final volatility of 60 percent is not unattainable. A manufacturer of aeration equipment for oxidation ditches (10) lists 55 percent volatility as attainable and as characteristic of "completely mineralized" sludge which does not require further stabilization. Efficient nitrogen removal by ENR goes hand in hand with substantial destruction of the volatile biomass generated during BOD removal. This should favorably affect the cost of sludge disposal.

The relationship between biomass destruction and change in sludge volatility may be expressed by the following two equations:

$$v = \frac{v_0(1 - b)}{1 - b\, v_0} \tag{8}$$

$$b = \frac{v_0 - v}{v_0(1 - v)} \tag{9}$$

where: v = final fractional sludge volatility
v_0 = initial fractional sludge volatility
b = fractional rate of biomass destruction.

The validity of these equations may be recognized by considering for instance that the inert mass of sludge containing 60 percent VSS was originally associated with 160 mg of VSS if v_0 was 0.8, as illustrated in the following example:

> In the course of an endogenous respiration procedure, 200 mg of sludge solids of 80 percent initial volatility ($v_0 = 0.80$) are reduced by 100 mg. Assuming that the entire weight reduction was with respect to volatile solids, the composition of the solids at the end of the procedure will be 40 mg fixed and 60 mg volatile ($v = 0.60$). The destruction of the volatile biomass amounts to 100 mg. This is (100/160) 100 = 62.5 percent biomass destruction. Using Equation 9 furnishes the same result, to wit:
>
> $$b = \frac{0.80 - 0.60}{0.80(1 - 0.60)} = 0.625 \text{ or } 62.5 \text{ percent}$$

In the operation of systems using ENR and aerobic sludge synthesis for N removal, two conditions must obviously be avoided or minimized: anoxic residence time in the absence of nitrates and loss of wastewater carbon in EOR. To minimize the occurrence of these conditions, it appears advisable to use a compartmentalized reactor, equipped with aeration and mixing equipment in such a way that the bulk of the compartments may easily be switched over from the aerobic to the anoxic condition and vice versa.

Two basic aeration patterns are conceivable, the block approach and the alternating zones approach, as shown in Figure 3. A compartmentalized reactor equipped with dual equipment (aeration and mixing) in most compartments would allow the use of either approach. In both approaches, the last cell should be an aerobic cell to nitrify ammonia nitrogen released in preceding cells and to strip gaseous nitrogen clinging to the floc.

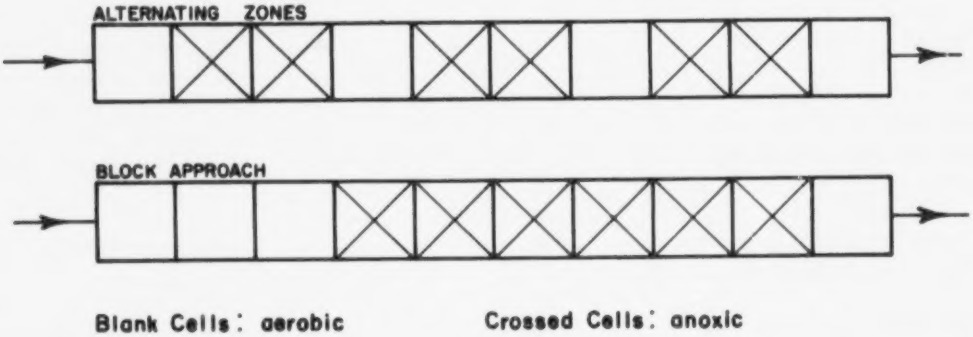

Figure 3. Aeration patterns in compartmentalized reactors.

If a reasonable number of cells are provided under the block approach, the daily aeration pattern could be modified in response to changes in load and reaction rates. When using the alternating zones approach, such adjustments might not be necessary.

Nitrogen Removal by Substrate Nitrate Respiration and Anoxic Sludge Synthesis

The stoichiometric equation for anoxic synthesis and substrate nitrate respiration combined in the ratio of 0.54:1 was listed as Equation 5. From that equation, the following table was abstracted.

TABLE 5. SUBSTRATE NITRATE RESPIRATION AND ANOXIC SYNTHESIS

Reduction of 1 mg of NO_3^--N is related to

Species	Action	mg
TOC	Removed	1.32
BOD_5	Removed	1.98
COD	Removed	4.40
Biomass	Produced	1.09
TKN	Metabolized	0.13
NH_4^+-N	Released	0.02
Alkalinity ($CaCO_3$)	Released	3.57

Approximately 2 mg of BOD_5 are needed for the gasification of 1 mg of NO_3^--N. This process of gasification also produces 1.09 mg of biomass, metabolizing thereby 0.13 mg of TKN. Whether or not NH_4^+-N is released in the reaction depends on the nitrogen content of the organic carbon source. Assuming that the organic carbon source has the composition $C_{10}H_{19}O_3N$, the net release of N to the process water (in the form of NH_4^+-N) is 0.02 mg per mg of NO_3^--N gasified. The biomass generated during substrate nitrate respiration and anoxic synthesis, obviously, could be used for nitrogen removal via an ENR anoxic reactor further downstream.

Substrate nitrate respiration is somewhat more difficult to implement in a single-sludge A.S. system than endogenous nitrate respiration. It becomes necessary to arrange for contact between nitrate nitrogen and untreated process water. But nitrates only become available after the process water has received a considerable measure of treatment. However, the carbon source in the substrate is removed from the process water after a very short length of treatment, this phenomenon being the basis of the contact stabilization process.

There seems to exist, basically, three ways to overcome these apparently contradictory requirements, to wit:

1) By step feeding the untreated process water to the reactor as illustrated in Figure 4. Only a portion of the carbon source in the wastewater is utilized for substrate nitrate respiration in this process, which is discussed at some length in Reference 5.

   The process water is divided into two or more portions. The first portion is fed, together with the return sludge, into the first aerobic cell of a compartmentalized aeration tank. The remaining portions are fed into anoxic cells further downstream. Interaction between the organic carbon source in these streams with nitrates generated in preceding aerobic cells brings about substrate nitrate respiration.

2) By mixed liquor recirculation, as suggested by Ludzak and Ettinger in 1962 (11). This procedure is now used in the first two reactors of the patented Bardenpho process (4) (12) (see Figure 5-21 of the EPA PDM). The stoichiometry of this process is also discussed in detail in Reference 5.

3) By pulsating aeration. The untreated process water is introduced into a basin which is alternately aerated and mixed anoxically. At the end of the aerobic period, the nitrogen of the process water will be mostly in the nitrate form. During the ensuing anoxic period, substrate nitrate respiration will occur via interaction between the inflowing untreated process water and these nitrates. However, the inflowing ammonia will remain nearly intact during the anoxic period. Nitrifiers are inactive at very low DO's. One way to prevent substantial bleedthrough of NH_4^+-N would be to place a nitrification reactor after the pulsating aeration basin. To obtain an effluent very low in NO_3^--N, some ENR will have to be provided as an additional step. For domestic wastewater, ENR does not occur in the pulsating aeration basin itself. During the anoxic periods,

the available NO_3^--N is completely utilized by the organic wastewater carbon source present. At the end of the anoxic period, the nitrates will be exhausted but some organic wastewater carbon source will still be available. In order for ENR to occur, NO_3^--N must be present but substrate must be absent. Several implementations of the pulsating aeration concept are discussed in the EPA PDM (see Figures 5-19 and 5-22 of that manual).

One cannot compare the nitrogen removal capacity (mg N removable/mg BOD_5 removed) of ENR with the removal capacity of substrate nitrate respiration per se. Removal capacities are comparable only in the framework of projected flow sheets. According to the third column of Table 4, the limit of the nitrogen removal capacity of ENR plus aerobic sludge synthesis is not far above 0.24 mg N/mg BOD removed. Reference 4 indicates that the Bardenpho flow sheet has approximately twice this nitrogen removal capacity. Under that flow sheet, the bulk of the nitrogen is removed via substrate nitrate respiration.

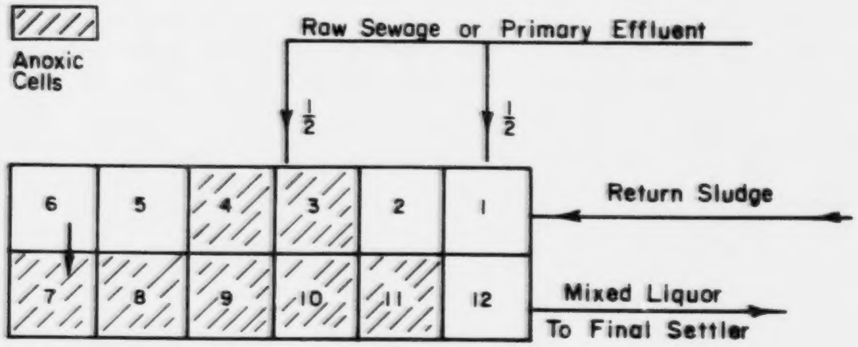

Figure 4. Two-step feeding of compartmentalized reactor.

NITRIFICATION KINETICS

In an A.S. system, the mass of the heterotrophic matrix per liter of reactor space (X, mg/ℓ) can be controlled directly by the operator by varying certain operational procedures, in particular the sludge recycle ratio (r) and the volume of the return sludge wasted per day (Q_w, ℓ/day). If the operator decides that his plant should be operated at a certain heterotrophic biomass concentration, X_a, he can implement this decision within fairly narrow limits, provided that adequate operating and measuring facilities have been provided.

For single-sludge A.S. systems designed for nitrification or nitrogen removal, the mass of nitrifiers per liter of reactor space ($\dot{N}$, mg/ℓ) cannot be determined with presently available analytical procedures in a practical way. The nitrifiers ($\dot{n}$.) are inseparably admixed to the heterotrophic matrix. The fractional content of the nitrifiers in this mixture varies. It follows that $\dot{N}$ cannot be controlled directly by the plant operator in the same way that he can control X.

As a measure of indirect control, the operator imposes a constant specific daily wasting rate on the n., thus causing them to vary in mass in response to load fluctuations and to assume a specific rate of growth which approaches, on the average, the imposed specific daily wasting rate. This fluctuating mass of n. would provide near constant quality effluent, i.e., NH_4^+-N concentration, if load fluctuations were sufficiently slow. By overdesigning the reactor with respect to average treatment requirements, effluent fluctuations under field conditions are kept within acceptable bounds. In the following paragraphs, "rate of growth" will be used frequently in place of "specific rate of growth." Also, the symbol X and the expression heterotrophic matrix (HM) will be used to designate the mixture of heterotrophic microbes and nitrifiers. The fraction of n. in the mixture (as estimated by research methods) is so small that X can be regarded as representing the heterotrophs only.

Nitrification in Completely Mixed (CM) Reactors

In CM reactors, the entire life span of the n. evolves under the influence of only one substrate concentration, the effluent NH_4^+-N concentration (S_1). Under otherwise constant environmental conditions, S_1 dictates the nitrifier growth rate (μ_n) in accordance with a rate of growth vs. substrate concentration function ($f_n(S)$) determined experimentally:

$$\mu_n = f_n(S) \tag{10}$$

The definitional equation for μ_n is:*

$$\mu_n = \left(\frac{1}{\dot{N}}\right)\left(\frac{d\dot{N}}{dt}\right) \text{, day -1} \tag{11}$$

This equation is valid for situations where the mass of nitrifiers is not affected by wasting.

The physical meaning of μ_n can be visualized by envisioning a CM reactor operated at constant $\dot{N}$, constant environmental conditions, and constant S_1 and utilizing a pure culture of n. $\dot{N}$ is kept constant by continuously wasting the biomass produced. The mass of n. wasted in one day per mass of n. present equals μ_n at the chosen S_1.

This case may be described by the following two equations:

$$\dot{N} = \dot{N}_0 e^{(\mu_n - w_n)t} \tag{12}$$

$$\mu_n = w_n \tag{13}$$

where:

$\dot{N}_0$ = n. concentration at time 0, mg/ℓ

$\dot{N}$ = n. concentration, mg/ℓ

w_n = specific n. wasting rate, day^{-1}.

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* The EPA PDM uses μ_n in a somewhat different sense.

The definitial equation for w_n is:

$$w_n = \left(\frac{1}{\bar{N}}\right)\left(\frac{d(\dot{n})}{dt}\right), \quad \text{day}^{-1} \qquad (14)$$

where: $d(\dot{n})$ = mass of n. wasted during dt per liter of reactor space, mg/ℓ.

Both μ_n and w_n are mass ratios referred to time. Equation 12 is also valid for those situations where $w_n \neq \mu_n$. Whenever $w_n \neq 0$, the definitial Equation 11 for μ_n becomes:

$$\mu_n = \frac{1}{\bar{N}}\left(\frac{d\dot{N}}{dt}\right) + w_n \qquad (11a)$$

The EPA PDM prescribes that the Monod function be regarded as the function (f_n). That recommendation is followed in this report. Equation 15, therefore, describes the functional relationship between μ_n and S:

$$\mu_n = f_n(S) = \frac{\mu_{max} S}{K_s + S} \qquad (15)$$

where: μ_{max} = maximum n. growth rate, in accordance with prevailing environmental conditions of wastewater temperature (T), pH, and DO, day^{-1} (μ_{max} is approached by μ_n at high S's, but it is never equalled by μ_n)

K_s = Michaelis-Menten constant, in accordance with prevailing T, mg/ℓ of NH_4^+-N.

The following values for μ_{max} and K_s are used in this report, in accordance with recommendations of the EPA PDM:

$$K_s = 10^{(0.051T - 1.158)}, \quad \text{mg/ℓ } NH_4^+\text{-N} \qquad (16)$$

$$\mu_{max} = 0.47\, e^{[0.098(T - 15)]}\left(\frac{DO}{1.3 + DO}\right)[1 - 0.8333(7.2 - pH)], \quad \text{day}^{-1} \qquad (17)$$

For a CM reactor, the functional relationship between μ_n and S_1 can, therefore, be described by either one of the following equations, provided that steady state conditions have been reached:

$$\mu_n = \mu_{nr} = \frac{\mu_{max} S_1}{K_s + S_1} \quad \text{(CM reactor)} \qquad (18)$$

$$S_1 = \frac{\mu_{nr} K_s}{\mu_{max} - \mu_{nr}} \quad \text{(CM reactor)} \qquad (19)$$

where: μ_{nr} = n. growth rate to be imposed on a reactor to satisfy given effluent requirements, day^{-1}.

In a CM reactor, the average daily growth rate equals the instantaneous rate of growth occurring anytime during the day, steady state conditions prevailing.

Equation 15 is plausible as growth is a consequence of biota-substrate interaction, and the rate of this reaction per mg of biota can be expected to increase with increasing substrate concentration. Equation 19 follows mathematically from Equation 18, but one can visualize quantitatively quite easily that it is indeed possible to produce a near constant effluent concentration in a CM reactor by imposing a constant rate of growth (μ_{nr}) on the biota therein, to wit: Assume that a steady state treatment situation has been reached. The rate of n. wasting per liter of sewage treated $(\Delta\dot{N}_w/\dot{N})_0$ equals the rate of n. production per liter of sewage treated $(\Delta\dot{N}_p/\dot{N})_0$ for any length of time. Now the influent NH_4^+-N concentration (S_0) is suddenly raised and kept at the increased value. $\dot{N}$ has to be assumed to stay unchanged, initially. S_1 will rise as illustrated in one of the following paragraphs. This will cause a rise in $\Delta\dot{N}_p/\dot{N}$ per Equation 12. The rate of n. production will now be larger than rate of n. wasting as the latter is kept constant by the operator. This leads to an increase in $\dot{N}$, which in turn will cause a decrease in $\dot{N}_p/\dot{N}$. $\dot{N}$ will increase until $\Delta\dot{N}_p/\dot{N}$ has reached again the initial value of $(\Delta\dot{N}_p/\dot{N})_0$, which is equal to $(\Delta\dot{N}_w/\dot{N})_0$.

If S_1 is determined only after steady state conditions have been reestablished following a change in load, it will be found that S_1 depends only on the μ_{nr} imposed on the reactor and on environmental conditions as defined by Equations 13, 14, and 16, but not on S_0, the rate of flow (Q), or the ratio of sludge recirculation (r). In the transition periods, S_1 will be higher than predicted by Equation 19 if the load was increased and lower if the load was decreased. The functional relationship between S_1 and μ_{nr} under steady state conditions is illustrated by Figure 5 for 15 C and 10 C. If the wastewater temperature falls while μ_{nr} stays constant, S_1 will increase.

Based on the assumption that $S_0 - S_1$ represents indeed the amount of NH_4^+-N nitrified in the process, $\dot{N}$ may be computed for a given S_0, sewage detention time (SDT_n), and μ_{nr} by using Equation 20:

$$\mu_{nr} = \frac{Y_n(S_0 - S_1)}{\dot{N}\ SDT_n} \tag{20}$$

where: Y_n = nitrification yield coefficient, in terms of NH_4^+-N

SDT_n = sewage detention time (nominal detention time) in the nitrification reactor, day.

Equation 20 is a translation of definitional Equation 11 into steady state conditions. $Y_n(S_0 - S_1)$ represents the mg of n. mass produced per liter of sewage treated. $1/SDT_n$ represents the liters of sewage treated per day per liter of reactor space. For all-aerobic systems, SDT_n may be replaced by SDT. According to the EPA PDM, Y_n has the value of 0.15.

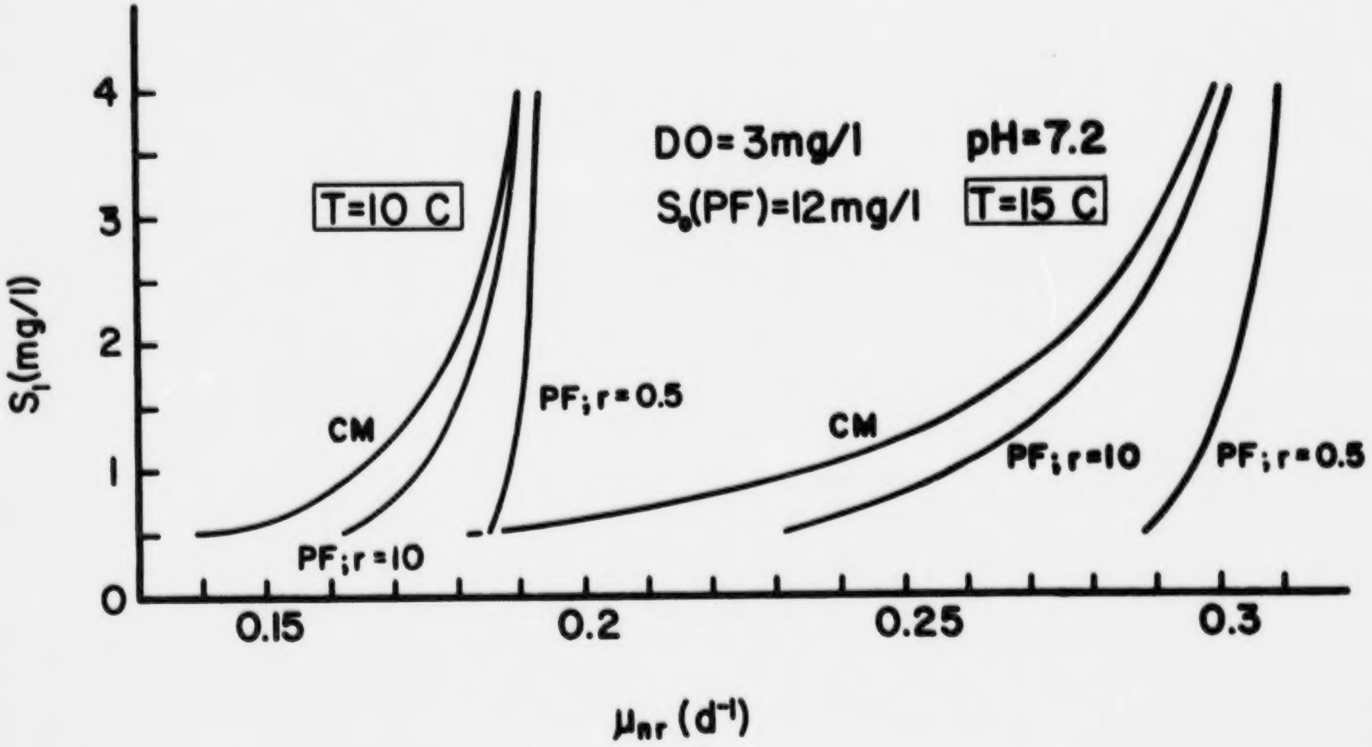

Figure 5. Effluent ammonia concentration (S_1) vs. imposed nitrifier growth rate (μ_{nr}).

The following definitial equations apply to SDT_n and SDT:

$$SDT_n = V_n/Q \tag{21}$$

$$SDT = V/Q \tag{22}$$

where: Q = plant flow or compartment flow, ℓ/day

V_n = capacity of nitrification reactor or compartment, ℓ

V = capacity of entire reactor in all-aerobic systems, ℓ.

For a treatment situation characterized by pH = 7.2, DO = 3 mg/ℓ, T = 15 C and S_1 = 1 mg/ℓ, a μ_{nr} of 0.2357 day^{-1} would be required. For S_0 = 14.5 mg/ℓ and SDT = 0.25 day, $\dot{N}$ would be 34.4 mg/ℓ. If the flow was increased to double the original value, SDT would be reduced to 0.125 day and $\dot{N}$ would increase to 68.8 mg/ℓ. If S_0 was increased to 29 mg/ℓ, $\dot{N}$ would increase to 71.3 mg/ℓ, again based on a SDT of 0.25 days. It should be noted however that a CM reactor operating at constant μ_{nr} cannot accommodate changes in environmental conditions of T, pH, and DO without a change in S_1.

Nitrification in Plug Flow (PF) Reactors

As a slug of MLSS travels through a PF reactor, the ammonia nitrogen concentration in the surrounding process water changes from a high of S_i (initial NH_4^+-N concentration after dilution of S_0 by the recycle flow) at the head end of the tank to a low of S_1 near the effluent end. μ_n varies in accordance with Equation 15 from a high of $\mu_{max} S_i/(K_s + S_i)$ to a low of $\mu_{max} S_1/(K_s + S_1)$. An average n. growth rate (μ_{na}) will result, assuming that all NH_4^+-N defined by $S_i - S_1$ is oxidized and that there is no addition to $S_i - S_1$ via ammonification:

$$\mu_{na} = \frac{Y_n (S_i - S_1)}{\dot{N}\, t_n} \tag{23}$$

where: t_n = flowthrough time (mixed liquor detention time) in the nitrification reactor, day.

By using the following obvious equalities:

$$S_i = \frac{S_0 + r\, S_1}{1 + r} \tag{24}$$

$$t_n = \frac{SDT_n}{1 + r} \tag{25}$$

$$\mu_{nr} = \mu_{na} \text{ (PF reactor)} \tag{26}$$

Equation 23, therefore, becomes identical with Equation 20. μ_{na} is the controllable growth rate of nitrifying reactors and forms the basis of kinetic equations.

A second approximating assumption is usually made when developing kinetics for PF reactors: that $\dot{N}$ stays constant during t_n. This simplification, made to facilitate the mathematics of the process, tends to understate the efficiency of the process by perhaps 3 to 5 percent.

For PF reactors, it is impossible to set forth a counterpiece to Equation 19. However, a counterpiece to Equation 18 may be written, to wit (13):

$$\mu_{nr} = \frac{\mu_{max}\,(S_0 - S_1)}{S_0 - S_1 + (1 + r)\,K_s \ln\left[\dfrac{S_0 + r\,S_1}{(1 + r)\,S_1}\right]} \quad \text{(PF reactor)} \qquad (27)$$

Derivation of this equation can be found in Appendix A.

Contrary to CM reactors, PF reactors are affected by changes in r and S_0. An increase in S_0 will bring about a decrease in S_1. An increase in r will bring about a deterioration of the effluent. Further aspects of the relationship expressed by Equation 27 can be recognized by observing Figure 5. For the PF reactors in this figure, an S_0 of 12 mg/ℓ was assumed.

That figure also provides a comparison between PF and CM, to wit:

a) CM generally requires a lower μ_{nr} than PF to achieve the same S_1; this means larger reactors with the CM regime.

b) The advantage of PF diminishes with an increasing prescribed S_1, with an increasing r used, and with a lower prevailing operating temperature.

c) PF is more sensitive to changes in μ_{nr} than CM.

Due to the higher prevailing average reaction rate (μ_{nr}/Y_n), PF reactors need somewhat less biomass than CM reactors for the same treatment situation.

The following data were computed for PF reactors. Corresponding data have been computed for CM reactors in the preceding subsection.

In order to provide an S_1 of 1 mg/ℓ at S_0 = 14.5 mg/ℓ, r = 0.5, T = 15 C, pH = 7.2, and DO = 3 mg/ℓ, it is necessary to provide a μ_{nr} of 0.30 day^{-1} for PF. For an SDT of 0.25 day, $\dot{N}$ will establish itself at 27 mg/ℓ. If S_0 is raised to 29 mg/ℓ, S_1 will be reduced to 0.2 mg/ℓ and $\dot{N}$ will increase to 38 mg/ℓ. If the flow is doubled at S_0 = 14.5 mg/ℓ, SDT will drop to 0.125 day and $\dot{N}$ will double from 27 to 54 mg/ℓ. These changes will occur, provided the n. biomass is given enough time to grow to the new $\dot{N}$ concentrations and μ_{nr} is maintained at 0.30 day^{-1}.

Imposing the Nitrifier Growth Rate

The plant operator imposes wasting rates on the HM, i.e., the MLSS. By doing so, he also imposes wasting rates on the mass of n.. The imposed n. wasting rates cause the n. to assume compatible growth rates over a period of time. Momentary n. growth rates cannot be controlled in CM reactors under conditions of fluctuating loads. In PF reactors, momentary n. growth rates cannot be controlled at all.

The identity of n. and MLSS wasting rates are based on the equation:

$$\Delta\dot{N}_w/\dot{N} = \Delta X_w/X \tag{28}$$

where: $\Delta\dot{N}_w$ = n. wasted per liter of sewage treated, mg/ℓ

ΔX_w = MLVSS wasted per liter of sewage treated, mg/ℓ

X = MLVSS concentration, mg/ℓ.

This equality holds because it can be assumed that the fraction of n. in the MLVSS is the same as in the waste sludge. Under steady state conditions, X is a constant.

To arrive at a convenient equation for μ_{nr} in terms of ΔX, X, and SDT, one makes use of the following obvious equalities which are valid at steady state conditions:

$$\Delta\dot{N}_p = Y_n\ (S_0 - S_1) \tag{29}$$

$$\Delta\dot{N}_p = \Delta\dot{N}_w \tag{30}$$

Combining Equations 20, 28, 29, and 30 furnishes the following equation for μ_{nr} in CM and PF reactors:

$$\mu_{nr} = \frac{\Delta X}{X\ SDT_n} \tag{31}$$

where: ΔX = MLVSS produced per liter of sewage treated, mg/ℓ.

Similarly, the growth rate imposed on the HM (μ_{hr}) is:

$$\mu_{hr} = \frac{\Delta X}{X\ SDT} \tag{32}$$

For all-aerobic reactors, μ_{nr} is the same as the growth rate imposed on the HM (μ_{hr}). For aerobic-anoxic reactors, the following relationship between μ_{nr} and μ_{hr} prevails, which follows readily from Equations 31 and 32.

$$\mu_{nr} = \mu_{hr}\left(\frac{SDT}{SDT_n}\right) \tag{33}$$

It must be emphasized that ΔX refers to the biomass production in the entire reactor, not only to the biomass production in the aerobic part. It will make no difference whether $\Delta X/X$ is based on MLSS or MLVSS.

Equation 31 may also be interpreted in the following way, which might look more plausible to some engineers. If f stands for the fraction of nitrifiers in the mixed liquor solids, f (ΔX) (Q) stands for the daily production of n. mass, and f (X) (V_n) stands for the mass of n. in the aerobic portion of the reactor; in order to obtain μ_n, the daily n. production must be related to the n. mass present in the aerobic portion of the reactor only. Nitrifiers present in the anoxic portion and in the settling tank must be regarded as dormant, not participating in growth. If they are not excluded from consideration, the values of the kinetic coefficients (Equations 16 and 17) could not be used. These coefficients were determined in all-aerobic reactors. Also, the definitional equation for μ_n (Equation 11) would not apply. Thus, it follows again:

$$\mu_{nr} = \frac{f\ \Delta X\ Q}{f\ X\ V_n} = \frac{\Delta X}{X\ SDT_n}$$

Safety Factor (SF) and Quantity of Nitrifier Biomass ($\dot{N}$)

In this report, SF shall mean the ratio of μ_{nr} to the design nitrifier growth rate (μ_{nd}) selected by the designer for imposition on the nitrification reactor, to wit:

$$SF = \frac{\mu_{nr}}{\mu_{nd}} \tag{34}$$

The application of an SF* means reducing μ_{nr} necessary to achieve a prescribed average S_1 under a given treatment situation to a smaller value. This will result in a lower S_1 for that treatment situation than prescribed.

For a PF reactor and a treatment situation characterized by S_0 = 14.5 mg/ℓ, r = 0.5, pH = 7.2, and DO = 3 mg/ℓ, a μ_{nr} of 0.30 will be required to achieve an S_1 of 1 mg/ℓ NH_4^+-N. Applying an SF of 2 will result in a μ_{nd} of 0.15 and a new S_1 of practically 0 mg/ℓ. For the same treatment situation and SF application, S_1 for a CM reactor will decrease from 1 mg/ℓ to 0.23 mg/ℓ as the growth rate is reduced from 0.24 day^{-1} to 0.12 day^{-1}.

Looking now at Equation 31, to wit:

$$\mu_{nr} = \frac{\Delta X}{X\ SDT_n}$$

it can be seen that there are three ways to reduce μ_{nr}, once all design parameters have been tentatively chosen:

a) Increase SDT_n by increasing the size of the nitrification reactor.

b) Decrease ΔX by measures taking effect outside of the nitrification reactor. Improving or providing primary settling or improving or providing ENR are two measures that might be taken. The first measure would reduce the

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* This SF concept differs from the SF concept of the EPA PDM as defined by Equation 3-29 on Page 3-20 of that manual.

BOD_5 to be removed in the activated sludge system, which in turn would decrease ΔX; the second measure would provide for a greater biomass destruction, also decreasing ΔX.

c) Increase X.

If either route b or c is taken, the application of an SF will cause an increase in $\dot{N}$ (from the value associated with μ_{nr}) by a factor slightly larger than the SF. This can be recognized by considering Equation 20, to wit:

$$\mu_{nr} = \frac{Y_n (S_0 - S_1)}{\dot{N} \; SDT_n}$$

The factor will be somewhat larger than the SF because a reduced μ_{nr} will also mean a reduced S_1 and, therefore, a larger $(S_0 - S_1)$ value. Applying an SF of 2 to the last described PF treatment situation will increase $\dot{N}$ from 27 to 58 mg/ℓ, not to 54 mg/ℓ. In a corresponding CM reactor, $\dot{N}$ would increase from 34 to 71 mg/ℓ.

The application of an SF amounts to a buffering of the reactor against sudden increases in load. This can be illustrated graphically for PF reactors as shown by the ammonia profiles in Figure 6.

Curve A is the NH_4^+-N profile for a reactor with an SDT of 0.25 day and an r of 0.5, treating a sewage with an S_0 of 14.5 mg/ℓ NH_4^+-N under environmental conditions that are characterized by $\mu_{max} = 0.33$ day^{-1} and $K_s = 0.4$ mg/ℓ. The effluent requirements call for $S_1 = 1.0$ mg/ℓ. Curve B is the NH_4^+-N profile for the same operating conditions but reflecting the effect of applying an SF of 2. Curve C reflects the effect of a sudden increase in S_0 from 14.5 to 29 mg/ℓ on the reactor that has produced Curve B. Applying tank scale II to Curve B furnishes ammonia Profile D, reflecting a sudden doubling of both the influent and recycle flow rates. Basic kinetic values for Profiles A to D are presented in Table 6. $\dot{N}$ is assumed to stay constant when the sudden load increase is applied.

TABLE 6. EFFECT ON BASIC KINETIC VALUES OF APPLYING SAFETY FACTOR OF 2 TO PLUG FLOW REACTOR

Profile	μ_{nr} (day^{-1})	S_o (mg/ℓ)	S_1 (mg/ℓ)	S_i (mg/ℓ)	$\dot{N}$ (mg/ℓ)	t_n (hr)	t_m* (hr)
A steady state	0.30	14.5	1	10.0	27	4	4
B steady state	0.15	14.5	0	9.7	58	4	1.8
C transient load	0.15	29.0	0.2	19.4	58	4	3.7
D transient load	0.15	14.5	0.5	9.8	58	2	1.8

* t_m = flowthrough time required to achieve prescribed S of 1-mg/ℓ.

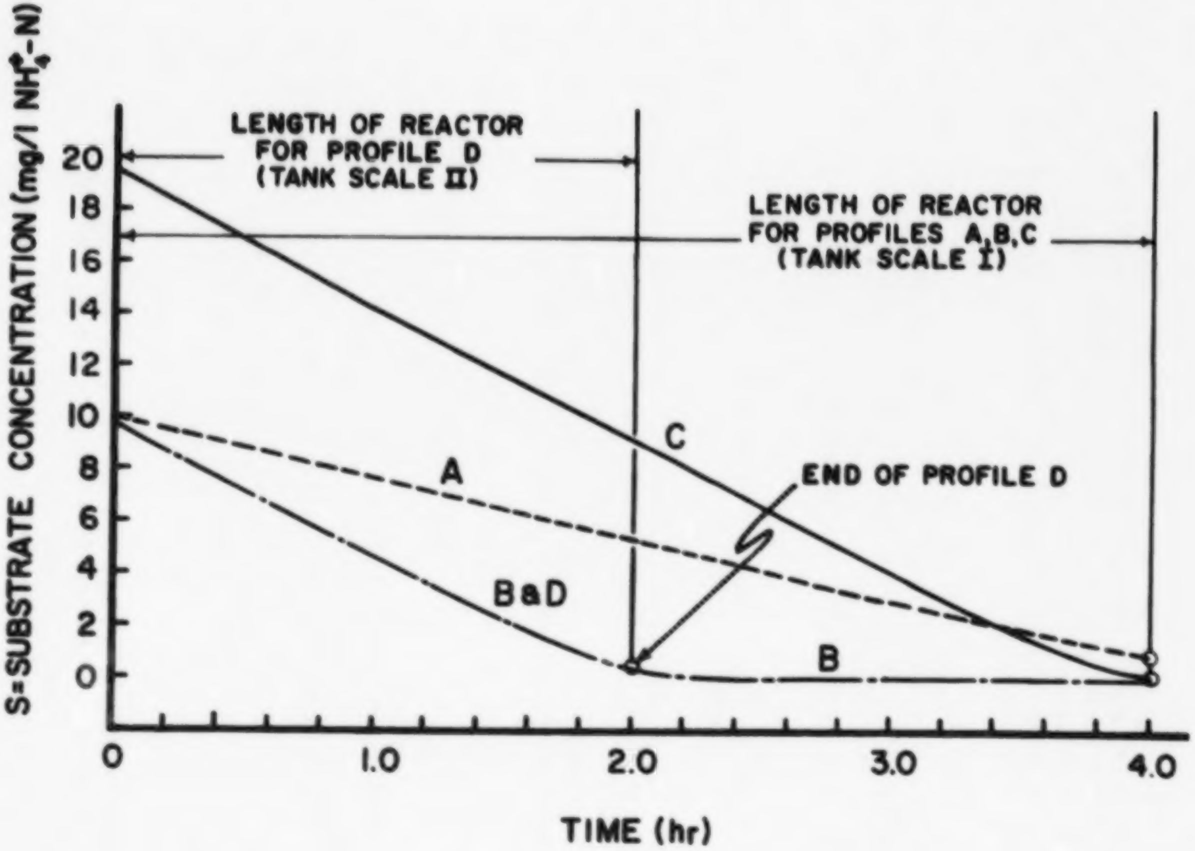

Figure 6. Ammonia profiles in a plug flow reactor illustrating Table 6.

Data points for Profiles A, B, and C were determined by evaluating Equation 35, which follows readily from Equations 23, 24, and 27.

$$t = \frac{Y_n}{\mu_{max}\,\dot{N}} \left[S_i - S + K_s \ln \left(\frac{S_i}{S} \right) \right] \tag{35}$$

where: S = the ammonia nitrogen concentration in the reactor at time t; note that S becomes S_1 for $t = t_n$. For Profiles A and B, $\dot{N}$ is the n. mass that has developed under the steady state operating conditions of these profiles. For Profiles C and D, $\dot{N}$ is the n. mass that has developed under the steady state conditions of Profile B.

$\dot{N}$ for Profile A was determined by using Equation 20; S_1 for Profile B was determined by using Equation 27 and for the computation of $\dot{N}$ for that profile, Equation 20 was used again. To determine S_1 for Profiles C and D, Equation 36 was used. In that equation, $\dot{N}$ is the n. mass developed under the steady state conditions of Profile B. Note that $SDT_n = (1 + r)\, t_n$:

$$SDT_n = \frac{Y_n}{\mu_{max}\,\dot{N}} \left[S_0 - S_1 + (1 + r)\, K_s \ln \left(\frac{S_0 + r\, S_1}{(1 + r)\, S_1} \right) \right] \tag{36}$$

Because of the slight difference in S_i (9.7 vs. 9.8 mg/ℓ), Curve B, read on tank scale II, represents Profile D only approximately.

To project the effect of a sudden increase in ammonia nitrogen load on a CM reactor, it is necessary to solve quadratic Fquation 37 for S_1:

$$S_1^2 + S_1 \left[\left(\frac{\mu_{max}\, SDT_n\, \dot{N}}{Y_n} \right) + K_s - S_0 \right] - S_0\, K_s = 0 \tag{37}$$

In this equation, which follows readily from Equations 18 and 20, $\dot{N}$ is the n. mass developed under the steady state conditions preceding the imposition of the transient load; the other parameters reflect transient load conditions.

With respect to buffering capability by application of an SF, CM reactors are somewhat less efficient than PF reactors. The following four values are obtained, respectively, for S_1 in a CM reactor for the four treatment situations of Profiles A, B, C, and D: 1.00, 0.23, 1.01, and 0.92 mg/ℓ NH_4^+-N.

The preceding discussion illustrates how the application of an SF will keep the effluent NH_4^+-N concentration below the prescribed steady state design requirement when a sudden increase in NH_4^+-N load occurs. If the increase in load is maintained for some time, the effluent NH_4^+-N concentration will start to decrease from the higher value experienced at the imposition of the transient load. This drop is in response to the increase in n. mass that will take place.

However, the effect of the SF on the effluent NH_4^+-N concentration as discussed is based on a simplification, an approximation to reality: it is based on the concept of a steady state load which does not exist in reality.

It should also be noted that the SF concept described does not fully correspond to the SF concept in other fields of engineering. Generally, the SF is the ratio of a selected design parameter value to the value of the same parameter computed in accordance to design requirements. If it were desired to use the same procedure for the computation of μ_{nd}, influent NH_4^+-N and BOD_5 loads would have to be known to the designer as mass vs. time functions and the required S_0, likewise, would be defined in the design requirements with respect to time and rate of flow, perhaps as the flow weighted average over 24 hr. From these data, μ_{nr} would be determined with the help of a computer program, taking fluctuations in n. mass during the design period into consideration. To the μ_{nr} thus determined, an SF would then be applied. In such a procedure, the SF would protect the plant against unforeseeable upsets. Load fluctuations would be accommodated in the determination of μ_{nr} itself.

CONCLUSIONS

This kinetic discussion of nitrification indicates that PF reactors are more efficient than CM reactors, particularly if a low effluent ammonia nitrogen concentration is desired and if sludge recycle rates can be kept low. Compartmentalization will minimize carbon loss via EOR in the warmer season if great fluctuations in wastewater temperature are expected. Compartments should be small at the effluent end of a PF reactor to minimize tankage operating at low ammonia nitrogen concentrations and, therefore, low reaction rates. Compartments at the head end of the reactors may be larger because the Monod function furnishes a near zero order reaction rate for NH_4^+-N concentrations above 2 to 3 mg/ℓ.

SECTION 5

ANALYTICAL METHODS AND SAMPLING PROCEDURES

ANALYTICAL METHODS

In general, Standard Methods (13th edition) (14) or methods recommended or approved by EPA (15) were used.

When determining dissolved phosphorus forms, 0.45 μm membrane filters were used. Paper towels were employed when determining the chlorine residual in the filtrate of chlorinated sludge. Glass fiber filters were used for all other dissolved parameters.

BOD samples were processed every day, 7 days a week. The same is true of SS and VSS samples. All other samples were processed during the 5 working days of the week. Refrigeration was employed for all samples that could not be processed immediately. Sulfuric acid was used to preserve samples collected on weekends and holidays that could not be processed on the day taken.

Chemical Oxygen Demand (COD)

In addition to the procedure described in Standard Methods, the autoclave method described by Albertson et al. (16) and a modified standard method were used. The autoclave method was used for 24-hr COD profiles of the final settler effluent.

On March 21, 1975, the standard COD method was modified by omitting the use of mercuric sulfate. This was prompted by a review of British standard methods (17) and by the difficulties encountered in obtaining $HgSO_4$ and disposing of laboratory wastes containing mercury. According to the British standards, the addition of $HgSO_4$ is not required for samples containing less than 100 mg/ℓ Cl^-. All the institutional sewage samples tested contained less than 50 mg/ℓ Cl^-.

Nitrite + Nitrate Nitrogen, Ammonia Nitrogen, and Orthophosphorus in Filtered Samples

These parameters were determined with a Technicon Autoanalyzer II using Technicon Industrial Methods Nos. 100-70W, 98-70W, and 94-70W, respectively (Technicon 1971).

Oxygen Uptake Rate (OUR)

OURs were determined immediately after sample collection. A 500-ml sample was brought to the laboratory in a stoppered 1000-ml fleaker (1), shaken for 30 sec, and poured into a 300-ml BOD bottle. The DO drop in mg/ℓ over 3 min was determined and furnished the basis for the determination of the OUR. A BOD bottle DO probe, equipped with an agitator, was utilized. OUR was referred to MLSS.

Total Organic Carbon (TOC)

TOC was determined on a Beckman Model 915 Total Organic Carbon Analyzer.

Alkalinity

Mixed liquor (ML) alkalinity was determined on filtered samples.

Filterability

The simple test described in the Culp-Culp textbook "Advanced Wastewater Treatment" (18) was used. The mℓ of filtrate produced in 30 sec were determined using a 9-cm (3.5-in.) Buchner funnel and No. 1 Whatman paper. The vacuum was kept at approximately 58 cm (23 in.).

Chlorine Residual

Chlorine residual on the filtrate of chlorinated sludge was determined by Standard Method 114A (Iodometric); all other chlorine residual determinations were made using Standard Method 114C (Orthodoline).

SAMPLING PROCEDURES

Two types of sampling methods were employed for collecting composite samples.

a) Samples were taken every hour and composited in accordance with flow. Refrigerated Serco samplers (2) were utilized for this method, which was employed for collecting raw sewage samples only.

b) Samples were taken every hour and composited as taken. The same amount of sample was collected every time. If the flowthrough rate was changed, a new compositing container was started. Shop-made, refrigerated samplers were used for this procedure, taking 170 ml of sample every hour.

Composite sampling was used for all main stream samples under Flow Sheets IV, V, and VI. For Flow Sheets I, II, and III, the mode of sampling used is indicated in the tables reporting the results of operations. ML and return sludge samples were taken as grab samples, twice daily.

In general, daily sampling was employed until the end of October 1974 including weekends. From then on, weekend sampling of raw sewage was discontinued. BOD_5 determinations on CCC effluent were also discontinued.

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(1) Product of Corning Glass, a capped beaker.
(2) Product of Sanford Products Corp., 2355 Rand Tower, Minneapolis, MN 55402

SECTION 6

TREATMENT FACILITIES, OPERATING PROCEDURES, AND PLANT MEASUREMENTS

TREATMENT FACILITIES AND OPERATING PROCEDURES

Pretreatment

The pretreatment units consist of two hand-raked screens in series. The first screen is equipped with bars of 25-mm (1-in.) spacing, the second with bars of 13-mm (1/2-in.) spacing. The original installation included two comminuters instead of the finer screens. The comminuters were taken out of service due to the high incidence of rags, pencils, and toothbrushes in the raw sewage. The comminuters could not cope with these materials.

Flow Equalization

The surge tank (Figure 7) is 9.76 m (32 ft) in diameter. Sludge accumulation at the bottom of the tank is prevented by either one of two procedures: (1) by operating a sludge scraper in conjunction with continuous recirculation of raw sewage or (2) by operating a floating aerator part of the time, also in conjunction with continuous recirculation of sewage.

The floating aerator is controlled by a time clock and equipped with a propeller-type impeller, coupled directly to a 7.45-kW (10-hp), 1150-rpm motor.

Operation of the sludge scraper results in somewhat better utilization of the surge tank space than provided by the floating aerator. When using the sludge scraper, the minimum liquid volume of the surge tank was 68.1 m^3 (18,000 gal); when using the floating aerator, the surge tank could be drawn down only to 90.8 m^3 (24,000 gal). The maximum liquid volume of the surge tank in both cases was 295.2 m^3 (78,000 gal).

Sewage is pumped from the surge tank by submerged, torque-flow type, constant-speed pumps to the sideweir structure shown in Figure 8. Due to the varying water level in the surge tank, the rate of pumping to the sideweir varies, with an average of 12.6 ℓ/sec (200 gpm). The sideweir returns part of the sewage to the surge tank and releases a constant but controllable flow to the plant. Flow to the plant from the sideweir is by gravity and is controlled by an adjustable submerged sluice gate opening.

Operation of a surge tank imposes a certain operational burden on the plant operator. A surge tank equalizes flow fluctuations that occur within a 24-hr period automatically, but it is obviously desirable to adjust the outpumping

1 . . . . . Influent
2 . . . . . Return Flow from Flow Control Structure (Sideweir)

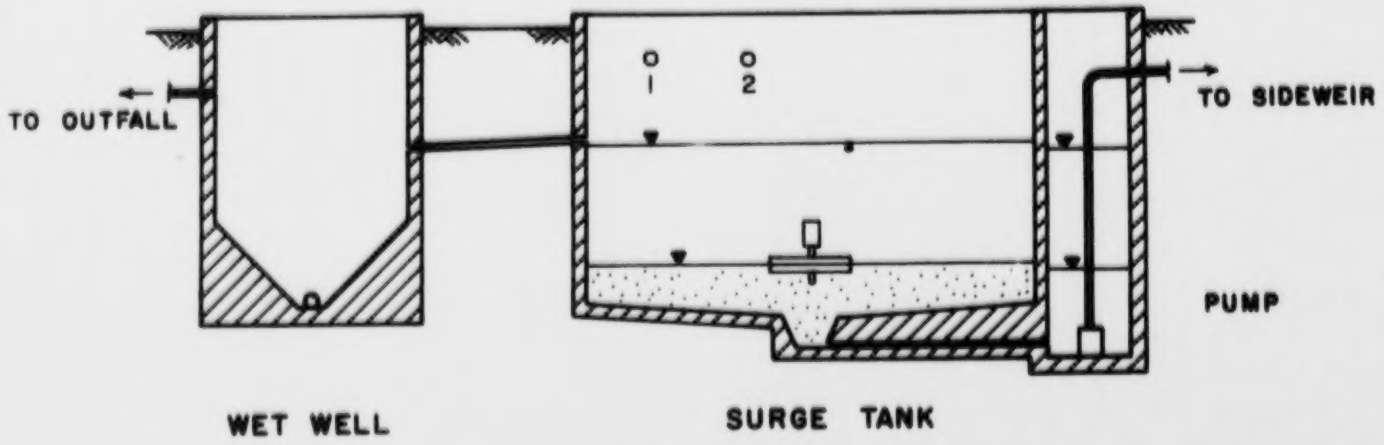
TO OUTFALL
1
2
TO SIDEWEIR
PUMP
WET WELL
SURGE TANK

Figure 7. Surge tank.

rate to fluctuations that occur from day-to-day. These nondiurnal fluctuations may be due to changes in water consumption from one weekday to the next, to changing rates of groundwater infiltration, or to changes of air temperature. At West Coxsackie, nondiurnal flow fluctuations are such that the highest 24-hr inflow rate to the plant is approximately double that of the lowest 24-hr inflow rate. The surge tank is equipped with two arrangements intended to compensate, if needed, for the selection of faulty outpumping rates: one is a low level pump cutout device, the other a bypass. The cutout device stops the pumps delivering sewage from the surge tank to the sideweir structure as soon as the fluid level in the surge tank falls below a certain level. The bypass allows settled and chlorinated sewage to be discharged to the receiving water whenever the overflow point is reached or in the event of a prolonged power outage. Settling and chlorination take place in the wet well shown in Figure 7. The wet well also serves as a backup or standby unit for the surge tank.

The net outpumping rate is controlled by manually varying the height of the sluicegate opening in the sideweir structure. Increments of 13 mm (1/2 in.) are used. The smallest sluicegate opening height used is 51 mm (2 in.), corresponding to a 4.5-ℓ/sec (71-gpm) plant flow. The largest opening is 114 mm (4-1/2 in.), corresponding to a plant flow of 9.1 ℓ/sec (144 gpm). At the higher flow rates, discharge from the submerged sluicegate is no longer free, i.e., the opening in the sluicegate becomes a submerged orifice.

Primary Settling Tank

The primary settler is of conventional, rectangular design, 1.83 m (6 ft) wide by 7.32 m (24 ft) long with a sidewater depth (SWD) of 2.44 m (8 ft). The sludge scraper-skimmer flights are chain operated. Sludge withdrawal is from a hopper at the head end of the tank. The sludge pump is a horizontal torque-flow pump, operated by a time clock. The original design provided a time clock allowing a minimum sludge pumping interval of 2 hr and a minimum period of pumping of 2 min. Only this arrangement was available for Flow Sheet I. Sludge pumping was at the rate of 12.6 ℓ/sec (200 gpm). For Flow Sheets V and VI (the only other flow sheets incorporating primary settling), adequate sludge pumping facilities were provided.

Aeration Tank

The aeration tank used is divided into 11 cells, numbered 1, 2, 3, 4, 5, 6, 7/8, 9, 10, 11, and 12 for the purpose of this presentation. Each cell, except Cell 7/8, is 1.52 m x 3.05 m (5 ft x 10 ft) in plan, with an SWD of approximately 2.3 m (7-1/2 ft) or 2.6 m (8-1/2 ft) depending on the height of the outfall round weir used. The 2.6-m (8-1/2 ft) SWD corresponds to 146 m^3 (38,500 gal) under aeration, the 2.3-m (7-1/2 ft) SWD to 129 m^3 (34,000 gal) under aeration (see Figure 9). Cell 7/8 is 3.05 m x 3.05 m (10 ft x 10 ft) in plan. Figure 2 illustrates the numbering scheme used.

Each cell, except Cell 7/8, is equipped with an INKA aeration grid of PVC pipe featuring 3.2-mm (1/8-in.) diameter air holes. Each grid covers 46 percent of the area of its cell. The grids are mounted 0.3 m (1 ft) from the bottom of the cells. They are attached to the floor of the aeration tank

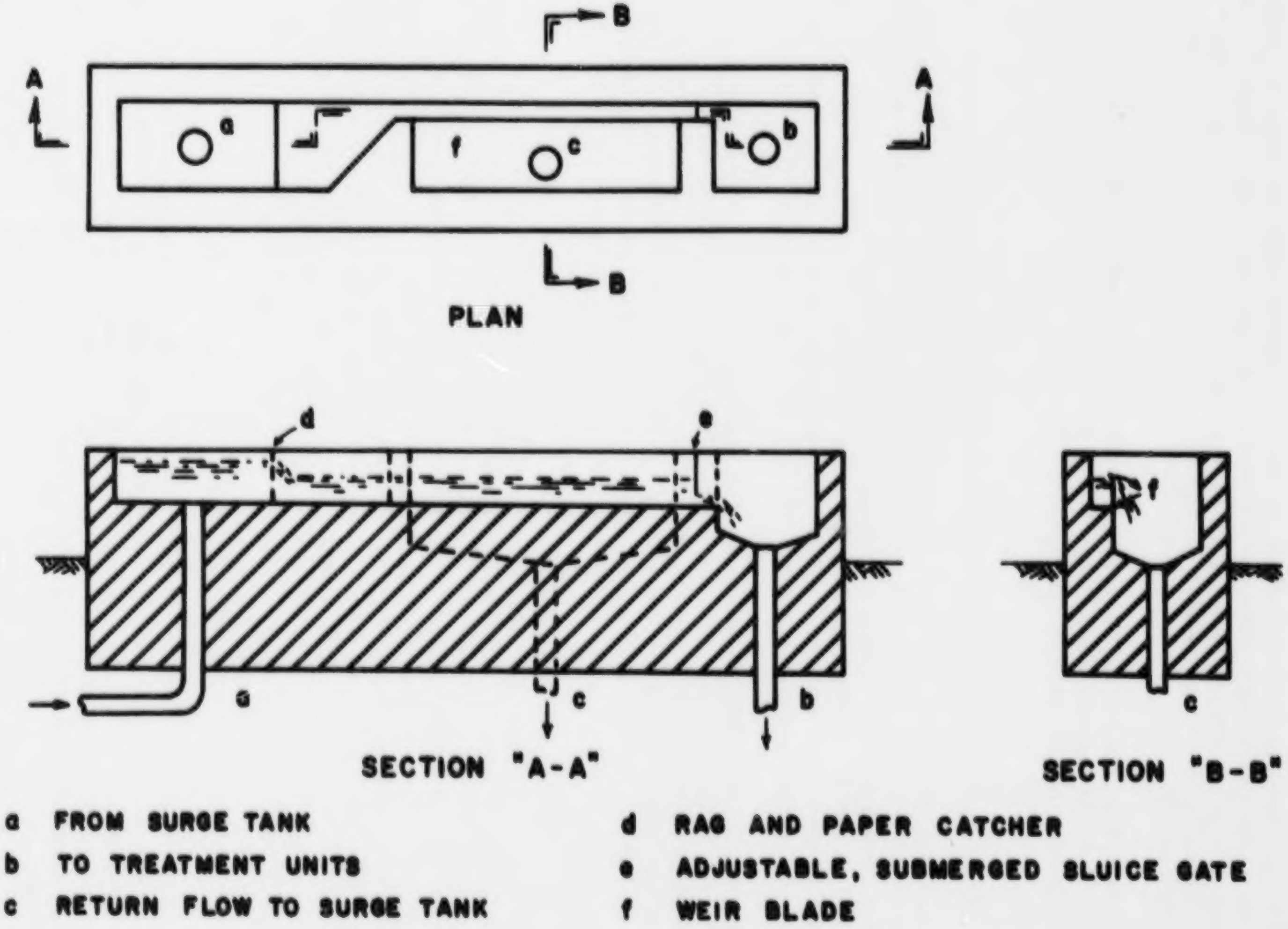
B
A
a
f
c
b
B
PLAN
d
e
a
c
b
f
c
SECTION "A-A"
SECTION "B-B"
a FROM SURGE TANK
b TO TREATMENT UNITS
c RETURN FLOW TO SURGE TANK
d RAG AND PAPER CATCHER
e ADJUSTABLE, SUBMERGED SLUICE GATE
f WEIR BLADE

Figure 8. Sideweir structure.

(Figure 9).

Cell 7/8 is equipped with a 0.25-kW (1/3-hp), 350-rpm mixer.* Anoxic conditions in other cells were achieved by throttling the air flow. Ports between the aeration cells are at the bottom of the partition walls and are simple slots 102 mm (4 in.) high by 305 mm (12 in.) wide. They are arranged in meandering fashion.

ML proceeds from Cell 1 to Cell 12 in PF fashion. However, there is a valved cross connection between Cell 5 and Cell 7/8 for use at higher flows. Both raw sewage and return sludge are admitted to Cell 1. ML is withdrawn from Cell 12.

Final Settling Tank, Sludge Recirculation, and Solids Management

The final settler is of conventional design with a 4.88-m (16-ft) diameter, a 3.05-m (10-ft) SWD, and peripheral-tangential entry of the ML. Tangential ML entry imparts a slow rotating motion to the entire sludge mass of the settler. A sludge blanket develops on the bottom of the settler and in the raceway between the skirt and peripheral wall; the blanket extends upward in the raceway to within approximately 0.4 m (1.3 ft) of the water level (Figure 10).

The height of the sludge blanket at the bottom of the settler was determined twice daily with the help of a Keene Model 8000 Sludge Blanket Finder. When the sludge blanket finder enters the sludge blanket, a slight modification of the audible signal occurs. The quality of this signal modification is indicative of the density of the sludge blanket and of the range of SS concentration in the effluent. A sharp change in tone indicates good flocculation and few SS.

For wasting of excess activated sludge (EAS) and for sludge recirculation, plunger pumps are used, the two pumps riding on the same suction line. The wasting pump is activated by a time clock at intervals of 2 hr. The quantity of sludge wasted on the project was varied by varying the sludge wasting periods; the shortest period used was 2 min every 2 hr. Occasionally, automatic wasting was replaced or supplemented by manual wasting. The amount of sludge wasted was governed by a general policy of maintaining a sludge blanket of 30-60 cm (1-2 ft) height.

Sludge Disposal

Two methods of sludge disposal are used:

a) Sludge chlorination to a pH of 2.5 followed by dewatering and drying of the chlorinated sludge on sand beds and disposal of the air-dried sludge on a landfill site. The sludge reacts with the chlorine for approximately 4 min in the pipeline that carries it from the point of chlorination to the sludge drying beds. This method is used during the months of May to October.

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* This corresponds to a specific power of 10 W/m^3.

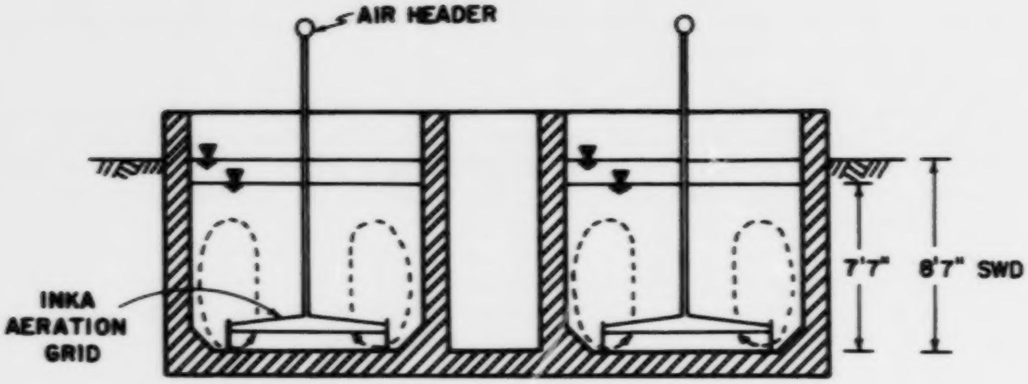
AIR HEADER
INKA AERATION GRID
7'7"
8'7" SWD

Figure 9. Section view of aeration tank.

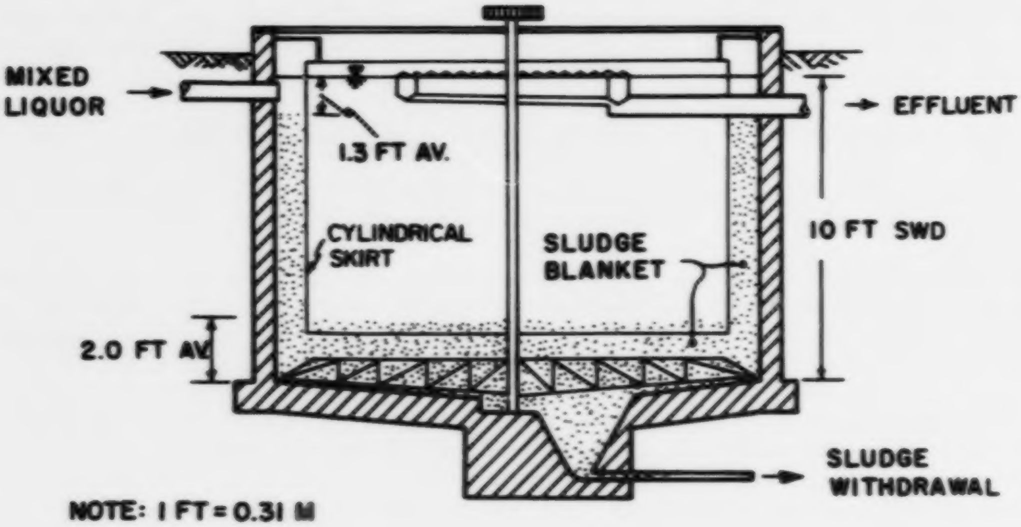
MIXED LIQUOR
EFFLUENT
1.3 FT AV.
CYLINDRICAL SKIRT
SLUDGE BLANKET
10 FT SWD
2.0 FT AV
SLUDGE WITHDRAWAL
NOTE: 1 FT = 0.31 M

Figure 10. Section view of final settler.

During the initial period of operations, a Purifax machine, manufactured by BIF, Providence, RI, was used for sludge chlorination. In early 1974, however, the reaction vessel of the machine was plugged by a rubber stopper and the pipeline leading to the sludge drying beds was employed as sole reactor from then on.

The total area of the sludge drying beds is 595 m^2 (6400 ft^2), composed of 149 m^2 (1600 ft^2) of standard sand drying beds previously used to dry anaerobically digested Imhoff tank sludge and 446 m^2 (4800 ft^2) of sludge drying area constructed on the top surfaces of two trickling filters, now out-of-service.*

b) Aerobic sludge stabilization and disposal of the treated sludge in a wooded area. This method is used during the remainder of the year, November to April. No supernatant recycle streams are directed back to the head end of the plant from the aerobic digester. The stabilized sludge is not decanted prior to disposal in the wooded area; however, sludge is thickened by gravity before being placed into the digester. When thickening EAS, a cationic polymer was used as a flocculation aid. Mixtures of EAS and primary sludge were thickened without such aid.

Ferric Chloride Feeding Procedure

Liquid ferric chloride solution of 42° Be' was purchased and fed as such with chemical feed pumps actuated by time clocks (percentage timers) on a 1-min or 2-min feeding cycle. Rates of feed were set in terms of mg/ℓ Fe^{3+} based on sewage flow through the submerged sluice gate in the sideweir structure. Rates of dosing, in mg/min, were adjusted concurrently with the sewage flow. The rate of ferric chloride addition was based on the treatment goal of obtaining an effluent phosphorus concentration of less than 1 mg/ℓ.

Chlorine Contact Chamber (CCC)/Second-Phase Settler Combination

The combined treatment unit, illustrated by Figure 11, consists of mixing chamber, influent plenum, flowthrough chamber, and effluent plenum.

In the CCC mixing chamber, chlorine solution is admixed to the final settler effluent. The chamber is equipped with a 0.25-kW (1/3 hp), 330-rpm mixer. The two plenums serve to distribute the process water over the cross section of the flowthrough chamber with the help of two orifice plates mounted at the entrance and exit of that chamber. Each orifice plate carries 40 orifices of 12.7 mm (1/2 in.) diameter. The flowthrough chamber is 1.52 m x 1.52 m (5 ft x 5 ft) in cross section and 4.88 m (16 ft) long.

On June 18, 1975, two baffle cages in series were installed in the flowthrough chamber. The baffle cages carry plexiglass baffles 3.2 mm (1/8 in.) thick that are inclined by 60° to the horizontal and spaced 152 mm (6 in.) on center or 305 mm (12 in.) vertically.

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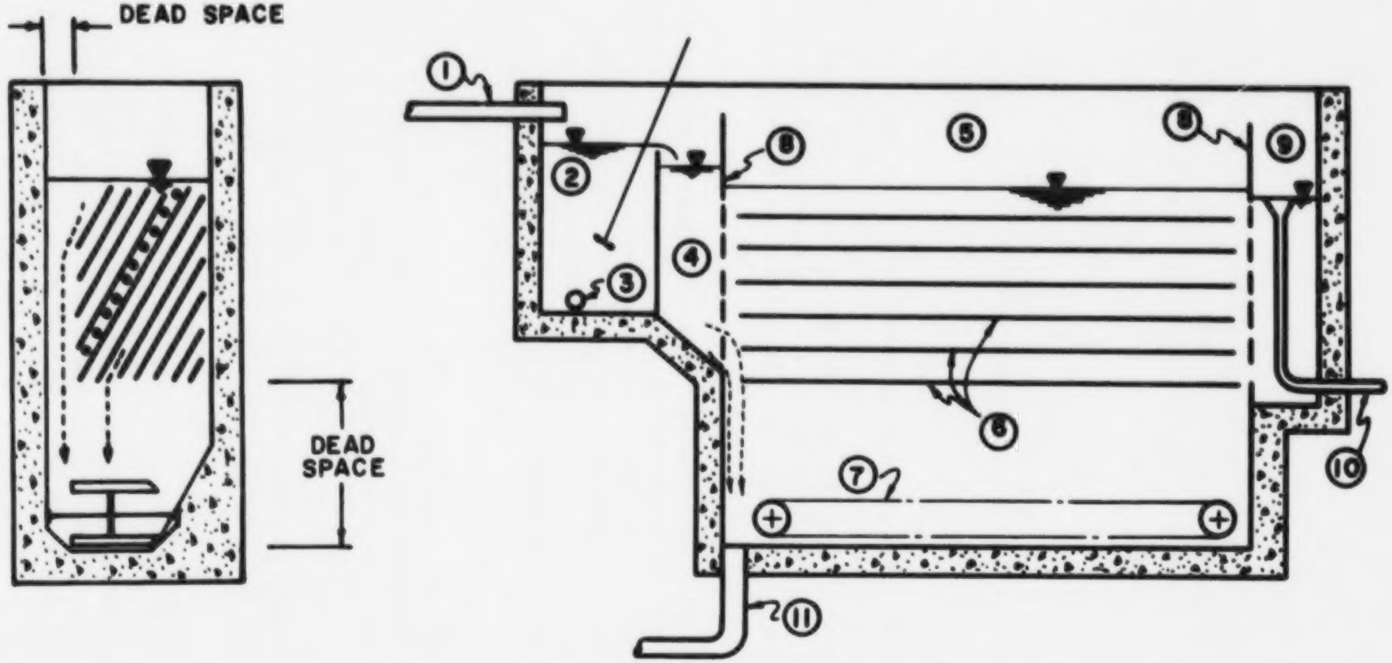
* Approximately 30 cm (12 in.) of stone was replaced by sand and gravel.


DEAD SPACE
DEAD SPACE
1
2
3
4
5
6
7
8
9
10
11

1. Secondary Settler Effluent
2. Mixing Chamber
3. Chlorine Solution Distributor
4. Influent Plenum
5. Settling Chamber
6. 60° Trays
7. Sludge Scraper
8. Orifice Plates
9. Effluent Plenum
10. Effluent
11. Sludge Withdrawal

Cross Section

Longitudinal Section

Figure 11. Chlorine contact chamber/second-phase settler combination.

The treatment unit is also equipped with a sludge scraper and sludge withdrawal means.

PLANT MEASUREMENTS

Surge Tank Detention Time

This parameter was determined by dividing the surge tank effluent flow into the average sewage volume of the surge tank, on a daily basis. The average daily sewage volume of the surge tank was determined from a level recorder chart.

Ferric Chloride Feed Rate

This chemical feed rate, expressed as mg/ℓ Fe^{3+}, was computed daily on the basis of a feed pump calibration curve and the surge tank effluent flow.

Main Stream Flow Measurements

The main stream flow was measured at two points, upstream of the surge tank and at the effluent point of the flow control device (sideweir structure) that controlled the flow to the treatment system downstream of the surge tank. The latter flow measurement, the surge tank effluent flow, was used for all computations in the report except for the arithmetic compositing of the raw sewage samples.

It should be noted that the reported raw sewage flows are approximately 20 percent higher than the surge tank effluent flows. The surge tank effluent flows were measured with a magnetic flow meter which was volumetrically calibrated with the help of stopwatch and the primary settler. The raw sewage flows were also measured with a magnetic flow meter. However, there was no possibility to calibrate that meter volumetrically.

Return Sludge Recycle Rate

Return sludge flow was measured with a magnetic flow meter which was calibrated volumetrically with the help of stopwatch and aerobic digester.

Aeration Tank Air Flow

Total air flow to the aeration tank was measured with an amperemeter. Air flow to the individual cells was not measured on this project. Ampere readings were taken daily.

Excess Activated Sludge (EAS) Production

The time-clock operated plunger pumps employed for wasting EAS were calibrated routinely. The volume of EAS per day was obtained by multiplying pumping time by delivery rate. EAS suspended solids concentrations were regarded as constant over 24-hr periods.

Aeration Cell DO

The DO level in the various cells was determined once or twice per day with a Yellow Springs DO meter.

SECTION 7

RAW SEWAGE CHARACTERISTICS

Although the Correctional Facility has a separate sanitary sewer system, infiltration occurs during rainy periods. In the fall, silo drainage (corn silage) enters the sewer system.

The institutional sewage is domestic in type but it is somewhat stronger, much fresher, and contains more rags than municipal wastewater. The raw sewage phosphorus concentration was somewhat higher during the study, approximately 4 mg/ℓ higher, than normally encountered in municipal sewage due to extensive use of an anticorrosion agent in the water treatment plant. ML temperature was also somewhat higher, approximately 7 C higher during the cold season, than in typical municipal sewage treatment plants, due to the very short sewer lines and the high water consumption in the institutional laundry. During winter, the sewage is cooled down several degrees during passage through the surge tank.

Annual average raw sewage and surge tank effluent characteristics for the 12 mo of August 1974 through July 1975 are presented in Table 7. Table 8 breaks down most of the same raw sewage characteristics on a monthly basis for the same 1-yr period. These data are based on 24-hr composites. The freshness of the sewage may be deduced from the comparatively low ammonia nitrogen and orthophosphorus content. Table 9 summarizes several raw sewage parameters for the fall of 1973.

During the cold season, the average influent flow was 5-10 percent lower than in the summer, with some increase in contaminant concentration (Table 8). The mass flow of contaminants stayed fairly constant. The major portion of the phosphorus in the raw sewage originates in the Correctional Facility's water treatment plant.

The raw sewage diurnal flow pattern is characterized by sudden changes in rate of flow, as shown in Figure 12. Minimum flow occurs in the early morning hours with three short peaks appearing throughout the day. Peaks I, II, and III reflect dishwashing and clean-up periods after meals; Peak III also reflects the use of showers. The instantaneous flow rate fluctuates between 2 ℓ/sec (32 gpm) and 25 ℓ/sec (396 gpm). The usage of showers affects the daily flow rate. Under the present liberal shower use policy, the daily flow rate increases considerably on hot weather days.

The iron content of the raw sewage fluctuated between 0.2 and 1.8 mg/ℓ. The higher values may have been caused by either sewage maintenance work or

TABLE 7. FLOW-WEIGHTED, ANNUAL-AVERAGE RAW SEWAGE AND SURGE TANK EFFLUENT CHARACTERISTICS FOR YEAR ENDING JULY 31, 1975 (FLOW SHEETS IV, V, AND VI)

Parameter	Raw Sewage	Surge Tank Effluent	Change	Change (% of Raw)	Comments
COD (mg/ℓ)	530	420	-110	-21%	
BOD_5 (mg/ℓ)	240	185	-55	-23%	
Diss. BOD_5 (mg/ℓ)	125	65	-60	-48%	
Diss. BOD_5 as % of BOD_5	52%	35%			
BOD_5 as % of COD	45%	44%			
SS (mg/ℓ)	233	243	+10	+4%	
VSS as % of TSS	84%				Sept., Oct., & Nov. 1974 only
Temp. of Sewage (C)	25.3	22.4	-2.9		
DO (mg/ℓ)	5.1	0.3	-4.8		
pH	7.4	7.4	0	0	
Alkalinity (mg/ℓ as $CaCO_3$)	175	200	+25	+14%	
NH_4^+-N (mg/ℓ)	9.7	12.7	+3.0	+31%	
TKN (mg/ℓ)	26.9	27.1	+0.2	+0.7%	(1)
NH_4^+-N as % of TKN	36%	47%			
Diss. Orthophosphorus (mg/ℓ)	4.4	4.8	+0.4	+9	
Diss. Phosphorus (mg/ℓ)	6.3	5.0	-1.3	-21%	9 mo. only
Total Phosphorus (mg/ℓ)	9.7	9.8	+0.1	+1%	(1)
COD/TKN	20	16			

(1) These increases appear to be caused by imperfect sampling or analytical error.

TABLE 8. FLOW-WEIGHTED, MONTHLY-AVERAGE RAW SEWAGE CHARACTERISTICS FOR 12 MONTHS ENDING JULY 31, 1975 (FLOW SHEETS IV, V, AND VI)

Flow Sheet / Parameter	IV Aug. 1974	IV Sept. 1974	IV Oct. 1974	IV Nov. 1974	IV Dec. 1974	IV Jan. 1975	IV Feb. 1975	IV Mar. 1975	V Apr. 1975	VI May 1975	VI June 1975	VI July 1975
Daily Flow (m^3)	606	689	568	629	606	591	602	583	580	580	648	712
Daily Flow (mil gal)	(0.160)	(0.182)	(0.154)	(0.166)	(0.160)	(0.156)	(0.159)	(0.154)	(0.153)	(0.153)	(0.171)	(0.188)
Temp. (C)	N.AV.	26.5	26.7	24.9	21.7	23.6	21.0	26.5	24.3	24.1	26.1	28.1
pH	N.AV.	7.2	7.5	7.7	7.7	7.4	7.5	7.3	7.3	7.3	7.3	7.2
Alk. (mg/ℓ as $CaCO_3$)	N.AV.	161	169	182	207	185	196	160	174	178	177	158
SS (mg/ℓ)	211	187	221	195	207	295	279	235	293	223	261	206
VSS (mg/ℓ)	N.AV.	129	202	176	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.
COD (mg/ℓ)	484	432	451	461	563	672	597	563	626	551	540	447
Diss. COD (mg/ℓ)	209	186	233	231	N.AV.	293	243	264	287	293	323	229
BOD_5 (mg/ℓ)	221	189	216	240	250	321	304	277	277	251	213	172
Diss. BOD_5 (mg/ℓ)	111	94	122	112	N.AV.	155	145	148	140	138	123	83
NH_4^+-N (mg/ℓ)	8.7	7.6	8.1	8.6	8.7	10.3	11.0	11.3	11.7	10.0	11.7	8.5
Org-N (mg/ℓ)	18.9	14.2	14.3	15.6	18.3	20.7	18.0	17.9	18.2	18.5	18.8	13.9
TKN (mg/ℓ)	27.5	21.9	22.1	24.2	26.9	31.0	28.5	29.9	29.9	30.1	30.5	22.4
NH_4^+-N as % of TKN	32	35	37	36	32	33	39	38	39	36	38	38
NO_2^-+NO_3^--N (mg/ℓ)	1.0	1.3	0.7	0.7	0.8	0.4	0.4	0.4	0.5	0.5	0.3	0.4
Total N (mg/ℓ)	28.5	23.2	22.9	24.9	27.7	31.5	29.5	30.2	30.3	30.6	31.4	22.9
TKN as % of COD	6	5	5	5	5	5	5	5	5	5	6	5
Ratio COD/TKN	18	20	20	19	21	22	21	19	21	18	18	20
Iron (mg/ℓ)	0.3	0.3	1.5	0.3	0.6	0.4	0.4	0.2	1.1	1.6	1.2	1.8
Diss. Orthophosphorus (mg/ℓ)	5.5	4.4	4.1	3.9	4.5	4.3	4.6	4.6	4.8	3.9	4.2	3.8
Diss. Phosphorus (mg/ℓ)	7.7	6.2	4.9	4.5	5.8	5.6	6.0	6.3	6.9	5.4	5.5	4.3
Total Phosphorus (mg/ℓ)	N.AV.	10.1	9.1	8.5	9.9	9.9	9.9	10.0	11.1	9.2	9.0	7.9
DO (mg/ℓ)	N.AV.	5.3	4.8	5.4	5.6	5.3	5.8	5.0	5.0	5.8	5.0	4.0

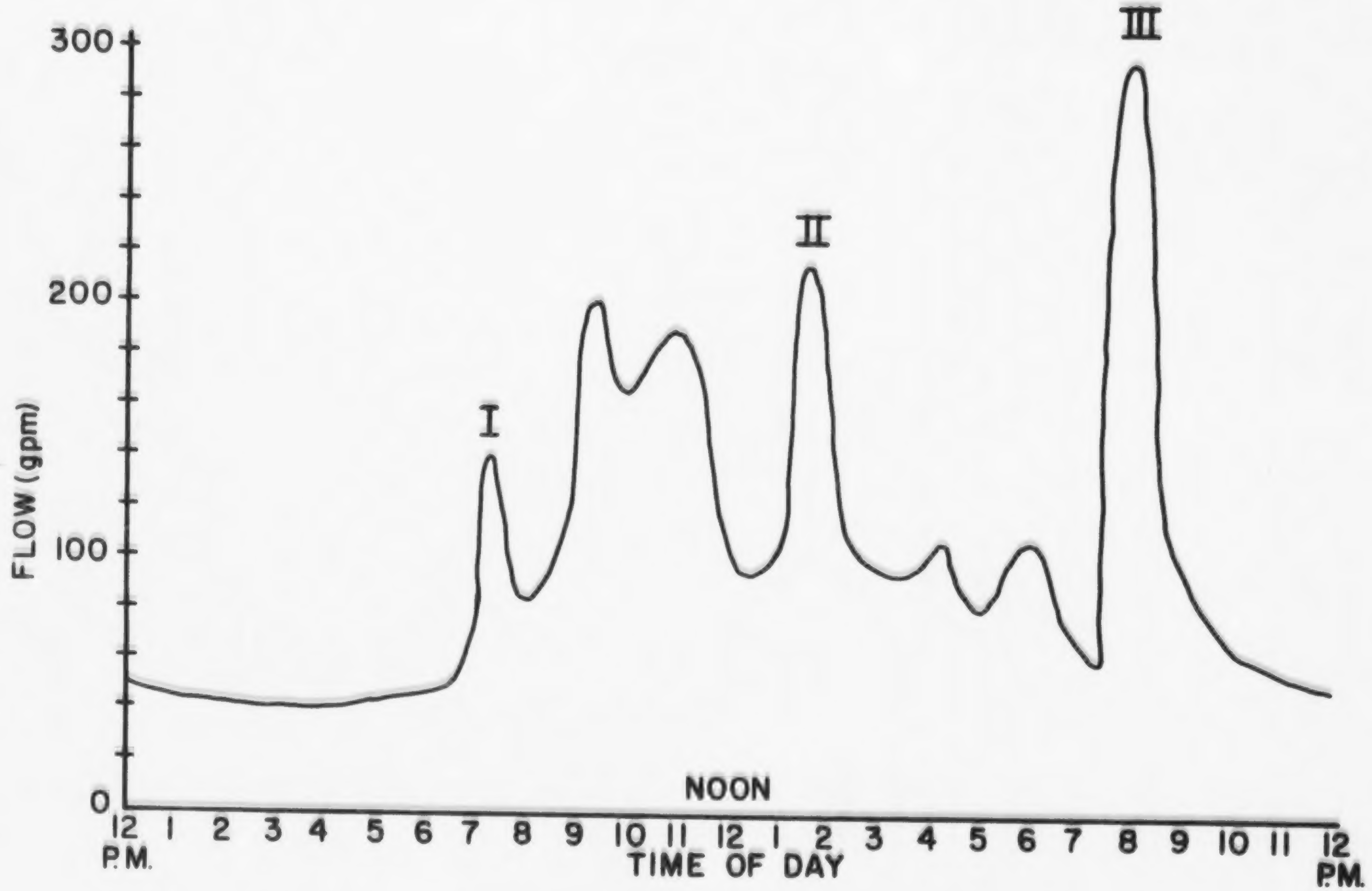


Figure 12. Raw sewage flow pattern (11/9/73 - 12/9/73).

TABLE 9. RAW SEWAGE CHARACTERISTICS FOR NOVEMBER AND DECEMBER 1973 (FLOW SHEET I)

Parameter	COD	BOD_5	SS	Total P	TKN	$NO_2^- + NO_3^-$-N
Number of Tests	6	3	5	2	5	6
Mean (mg/ℓ)	470	233	211	9.5	28	0.1

farming operations. The raw sewage contains minimal grit, but it contains a noticeable amount of Anthrafilt, originating in the water filtration plant. When backwashing the multimedia filter, some Anthrafilt is carried along with the washwater. The carriage water used by the Correctional Facility is surface water, drawn by gravity from an impounding reservoir located nearby. At times, the Anthrafilt in the raw sewage caused serious clogging problems in the sewage treatment plant.

During dry weather, all the nitrogen contained in the raw sewage is present as organically bound nitrogen and ammonia. At times of heavy infiltration, the nitrate-nitrite nitrogen content of the influent rises to 3-4 mg/ℓ.

SECTION 8

RESULTS AND DISCUSSION

OPERATING RESULTS: FLOW SHEET I

This flow sheet, as shown schematically in Figure 13, was characterized by two anoxic zones in the aeration tank, by the incorporation of the primary settler in the flow regime, by the anoxic operation of the surge tank, and by ferric chloride addition to the effluent of the primary settler. The raw sewage flow was somewhat smaller than experienced under subsequent flow sheets due to a smaller inmate population.

The sampling technique available during Flow Sheet I operations was poor, especially during cold weather when many sampler breakdowns occurred. Also, removal efficiencies for the primary settler were quite low for this phase due to an unsatisfactory mode of wasting primary sludge. On the other hand, the low average influent sewage flow and the anoxic detention time provided in the treatment train seemed to affect treatment results favorably, erratic as they may appear. The erratic results achieved under this flow sheet were likely due not to an unreliable process, but rather to poor operation and process monitoring.

Results achieved and operating conditions prevailing under Flow Sheet I are reported in Tables 10 and 11. The nitrogen removal aspects of this flow sheet are discussed in the subsection on Nitrogen Profiles, beginning on page 95.

Surge Tank Operation

Detention time in the surge tank was approximately 9 hr. Because the surge tank aerator was not utilized during this phase, a slight sewage odor was always noticeable near the surge tank. Considerable floating matter was visible in the surge tank. The color of the sewage in the surge tank and in the primary settler was milkish white.

Primary Settler Operation

Detention time in the primary settler was approximately 1.7 hr. Due to faults of design, there was always some floating sludge from the sludge hopper on the settler surface.

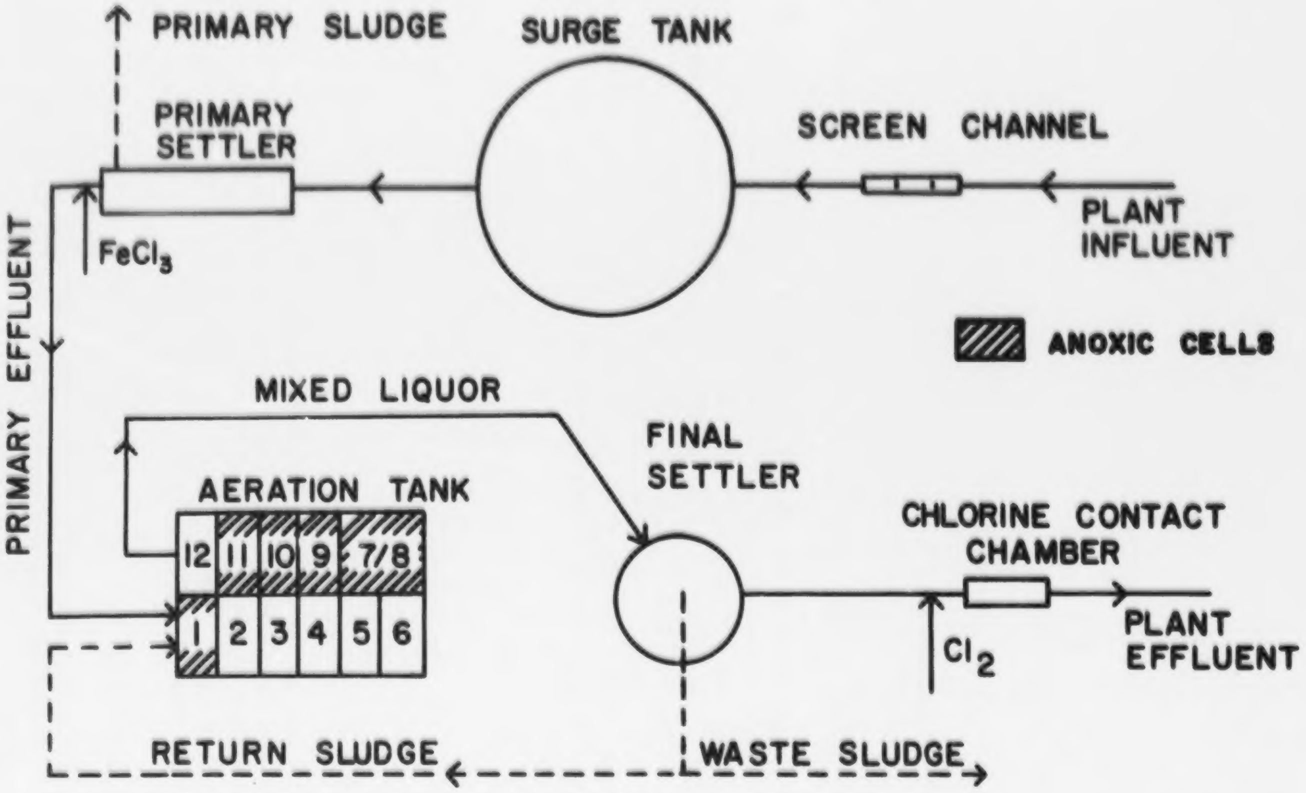

PRIMARY SLUDGE
SURGE TANK
SCREEN CHANNEL
PRIMARY SETTLER
PLANT INFLUENT
FeCl$_3$
ANOXIC CELLS
PRIMARY EFFLUENT
MIXED LIQUOR
FINAL SETTLER
AERATION TANK
12
11
10
9
7/8
1
2
3
4
5
6
CHLORINE CONTACT CHAMBER
PLANT EFFLUENT
Cl$_2$
RETURN SLUDGE
WASTE SLUDGE

Figure 13. Process schematic for Flow Sheets I and II.

TABLE 10. AVERAGE VALUES OF OPERATING PARAMETERS FOR FLOW SHEET I

Parameter	December 1973	January 1974	February 1974	March 1974
Sewage Flow (mgd)	(0.122)	(0.130)	(0.124)	(0.127)
Sewage Flow (m^3/day)	461	492	469	481
Detention Times Based on Raw Sewage Flow				
Surge Tank (hr)	11	7	9	10
Primary Clarifier (hr)	1.7	1.6	1.7	1.6
Secondary Clarifier (hr)	3.0	2.8	2.9	2.9
Aeration Tank (hr)	6.7	6.3	6.6	6.5
Detention Time Based on Mixed Liquor Flow				
Aeration Tank (hr)	5.0	4.8	5.0	4.9
Loadings and Overflow Rate				
*Organic Loading Rate, Aeration Tank - F/M (kg BOD_5/day/kg MLVSS)	0.22	0.25	0.28	0.29
*Volumetric Loading Rate, Aeration Tank				
(kg BOD_5/day/m^3)	0.56	0.68	0.66	0.68
(lb BOD_5/day/1000 ft^3)	(35)	(42)	(41)	(42)
Overflow Rate, Final Settler				
(m^3/m^2/day)	24.8	26.4	25.1	25.7
(gpd/ft^2)	(609)	(648)	(615)	(629)
Mixed Liquor and Sludge Production Data				
Sludge Retention Time (days)	5.8	6.3	4.1	4.9
kg VSS produced per kg BOD removed	0.75	0.61	0.84	0.68
MLSS (mg/ℓ)	3900	3900	3300	3400
Volatility [(MLVSS/MLSS) X 100]	64%	70%	71%	68%
Rate of Sludge Return	33%	30%	31%	33%

* Loading rates are based on kg of BOD_5 introduced into the aeration system.

TABLE 11. AVERAGE CHLORINE CONTACT CHAMBER EFFLUENT VALUES FOR FLOW SHEET I[†]

Parameter	December 1973	January 1974	February 1974	March 1974
pH	7.3	7.0	7.0	7.0
(Number of Observations)	(8)	(21)	(17)	(19)
BOD_5(mg/ℓ)[*]	3.0	3.8	4.8	5.8
(Number of Grab Samples)	(6)	(18)	(13)	(7)
SS (mg/ℓ)	4.0	7.4	5.5	6.3
(Number of Grab Samples)	(15)	(8)	(17)	(11)
TKN (mg/ℓ)	1.38	1.55	3.36	3.45
(Number of Grab Samples)	(6)	(14)	(17)	(4)
NO_2^-+NO_3^--N (mg/ℓ)	2.1	2.1	0.6	2.1
(Number of Grab Samples)	(9)	(17)	(17)	(16)
Soluble Orthophosphorus (mg/ℓ P)	0.16	0.33	1.09	0.25
(Number of Grab Samples)	(9)	(17)	(17)	(17)

[†] Average chlorine dose was 9.5 mg/ℓ.

[*] BOD_5 samples were seeded before analysis.

Aeration Train Operation

Nominal detention time averaged approximately 6.5 hr, r 32 percent, and MLSS concentration 3500 mg/ℓ.

Aeration Train Profiles

On November 14, November 27, and December 5, 1973, profile sampling was conducted through the aeration train. Samples were drawn from the individual cells in sequence, spacing the sample taking at intervals corresponding to an ML flowthrough time of 29 min per cell (58 min for Cell 7/8). The resulting profiles are shown in Figures 14, 15, 23, 24, and 25. Zero time samples were not collected. The first samples were drawn from Cell 1, representing the effluent from that cell and the influent of Cell 2. For tabulation of profile data, see Appendix B.

Pattern of Aeration

The first cell was kept anoxic because it was felt that such an arrangement would contribute to keeping the sludge volume index (SVI) down, with a view to the work of Chudoba et al. (19). On the basis of their laboratory investigation, Chudoba and his coworkers came to the conclusion that a steep

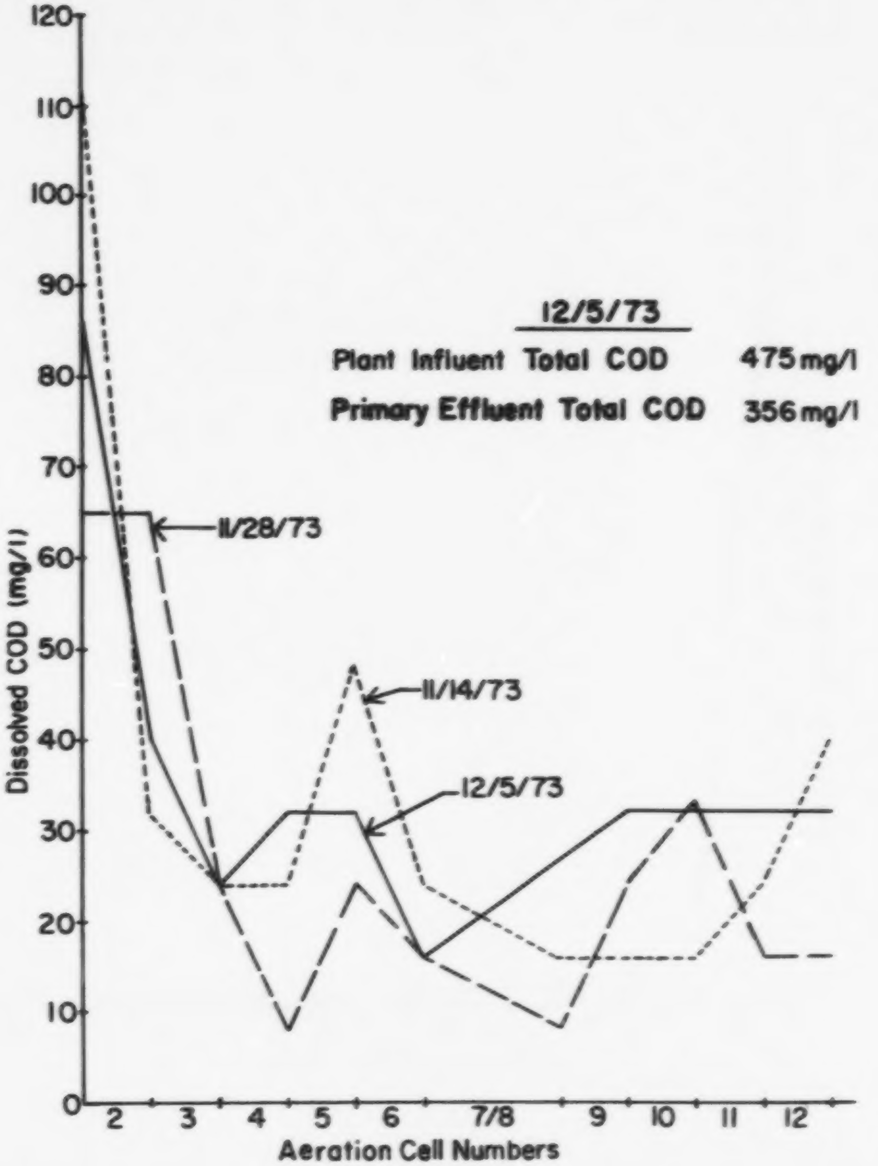


Figure 14. Aeration train dissolved COD profiles for Flow Sheet I.

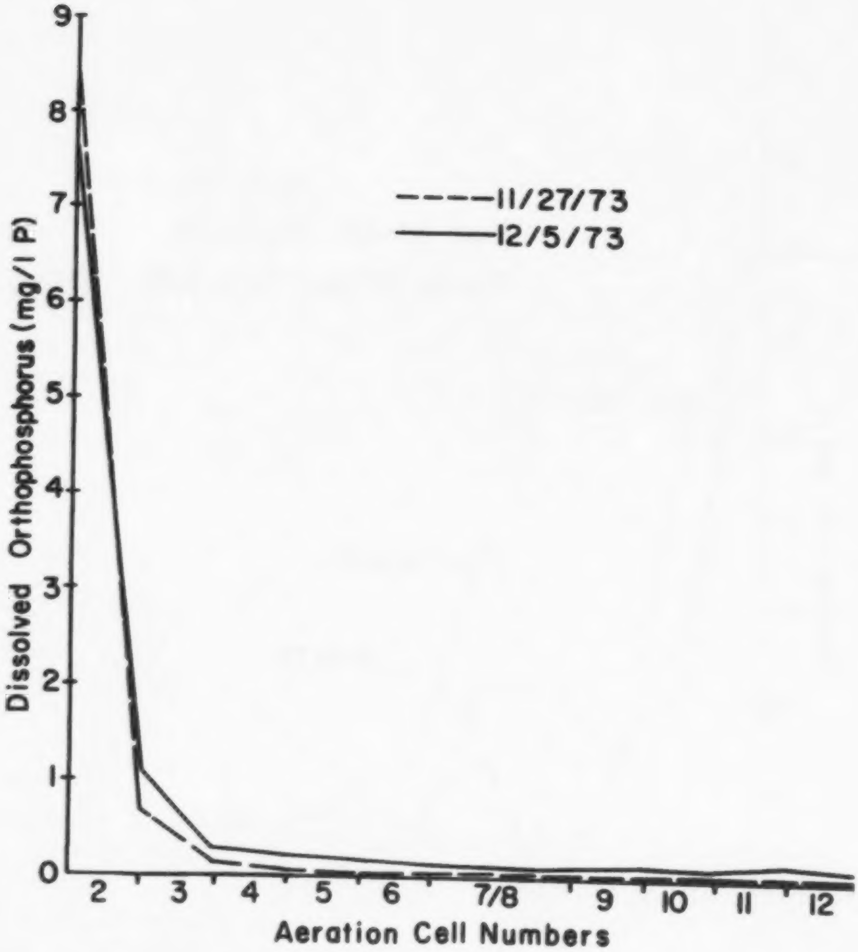


Figure 15. Aeration train dissolved orthophosphorus profiles for Flow Sheet I.

organic carbon gradient at the head end of the aeration tank would discourage sludge bulking due to filamentous organisms.

Final Settler Effluent Profiles

Total COD and NO_3^--N 24-hr profiles were run on the final settler effluent for 3 days in November and December of 1973. These are depicted in Figures 16 and 17.

Carbon Removal

The profile tables (Tables B-1, B-2, and B-3 in Appendix B) and Figure 14 indicate definite trends for the first half of the aeration train:

1) Nearly all of the particulate COD in the primary effluent is adsorbed in the first cell, the anoxic cell, by the MLSS. This is indicated by the low SS concentration in the supernatant of the first cell ML (see Tables B-1, B-2, and B-3). For approximate SS content of primary effluent, see Table 32.

2) A very rapid reduction of soluble COD takes place in the upstream portion of the aeration tank, consisting of the anoxic cell and the first two to four aerobic cells. This reduction brings the soluble COD concentration of the process water down to practically the final effluent soluble COD concentration.

3) After the original rapid reduction of soluble CODs in the aeration train, an increase in this parameter occurred, indicating perhaps release of organics from the ML biomass. On two of the three occasions depicted in Figure 14, subsequent anoxic treatment in double Cell 7/8 resulted in resorption of the released COD, but the data for the following anoxic cells are somewhat diffuse.

Soluble Orthophosphorus Removal

The aeration train profile shown in Figure 15 indicates that approximately 7 mg of soluble orthophosphorus were adsorbed to or absorbed by the ML sludge in Cell 2, the first aerobic cell. This shows very clearly that the aerobic portion of the activated sludge process itself was responsible for phosphorus removal. It should be noted that the activated sludge was in full contact with soluble orthophosphorus in Cell 1; however, no phosphorus removal occurred there. The bulk of the remaining soluble orthophosphorus was removed in Cell 3. Of further noteworthiness was the lack of any phosphorus release in anoxic Cells 7/8, 9, 10, and 11.

The mechanisms of phosphorus removal via $FeCl_3$ addition did not become clear under this research project. Total phosphorus in the raw sewage averaged approximately 10 mg/ℓ. During the time the aeration tank profiles were taken, only 13 mg/ℓ of Fe^{3+} were needed to achieve nearly complete transfer of the phosphorus to the sludge. Stoichiometric requirements were 18 mg/ℓ. Removal of some particulate phosphorus with the primary sludge and biological

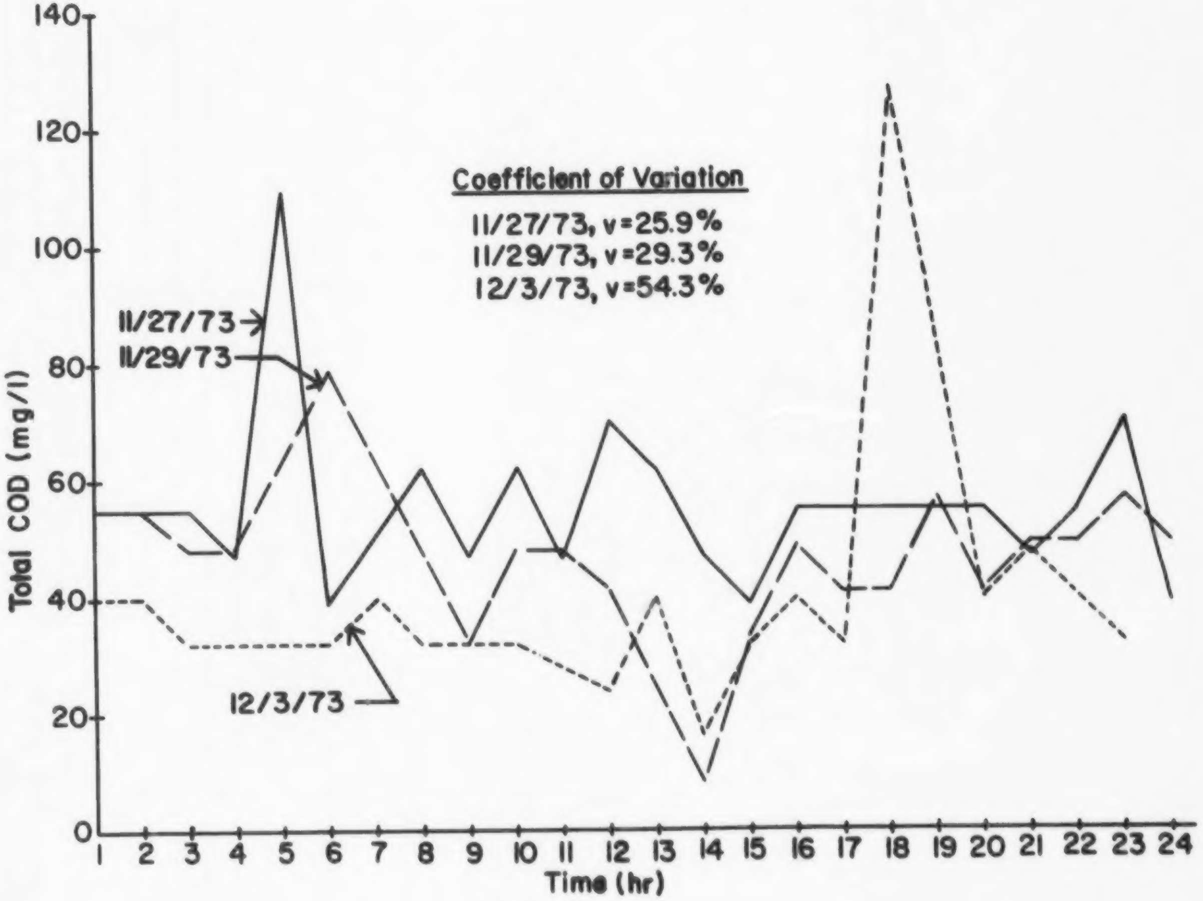


Figure 16. Daily final settler effluent total COD profiles for Flow Sheet I.

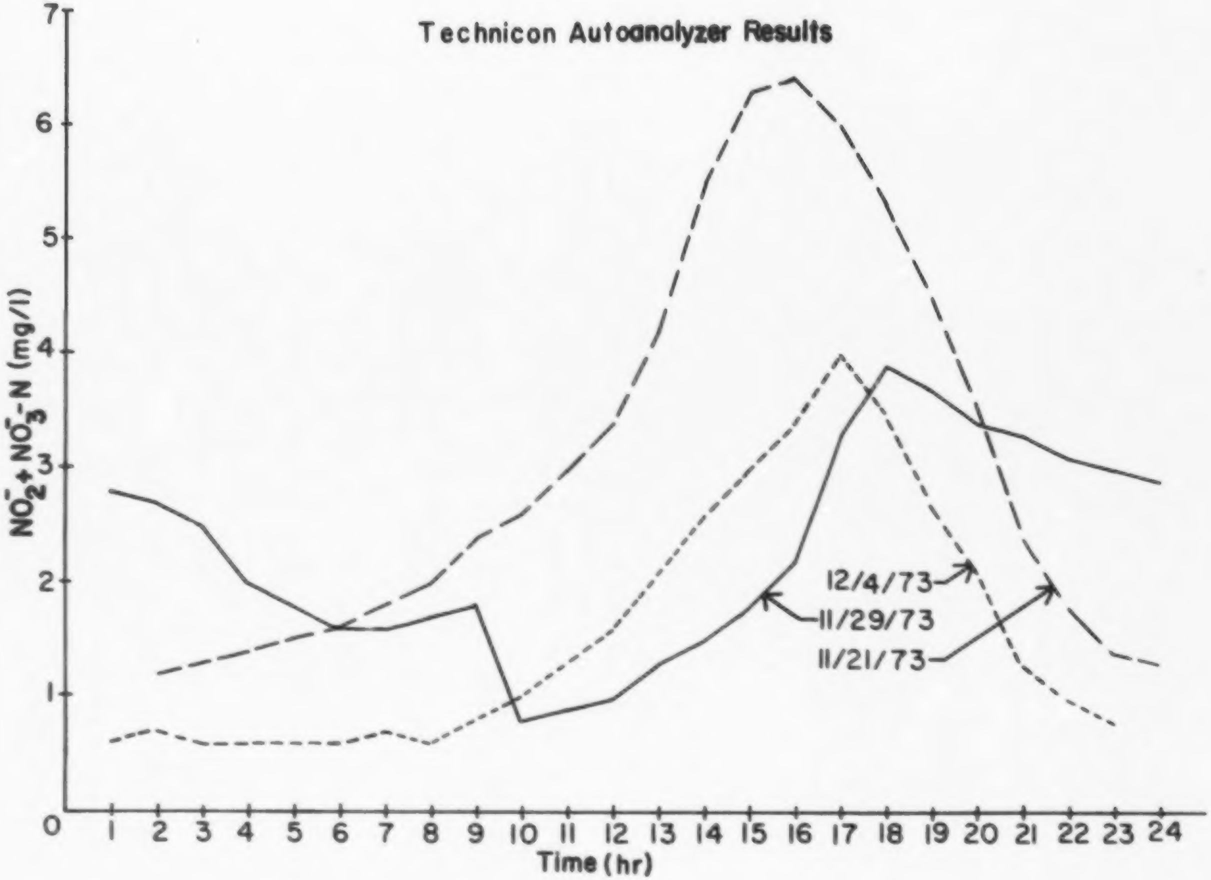


Figure 17. Daily final settler effluent NO_2^-+NO_3^--N profiles for Flow Sheet I.

phosphorus removal in the aeration tank to satisfy synthesis requirements may account for the efficient performance at lower-than-stoichiometric iron doseages. Also, the anoxic conditions in the first aeration cell and/or in the surge tank may have created a treatment situation favorable to luxury uptake of phosphorus.

The ML sludge underwent a visible color change with rising ferric chloride dosing rate, becoming darker and reddish-brown, indicating an accumulation of ferric hydroxide or hydrated ferric oxide. Also, when standing anaerobically in a glass bottle in the laboratory, the sludge turned black within a few days. This suggests formation of ferrous sulfide as the result of a reaction between ferric hydroxide and hydrogen sulfide.

Final Effluent Oxidized Nitrogen Profiles

Figure 17 depicts three 24-hr $NO_3^-+NO_2^-$-N profiles of the final settler effluent. These around-the-clock cross sections indicate a very pronounced oxidized nitrogen peak at the late afternoon hour, probably reflecting a morning nitrogen peak load on the plant due to kitchen and bathroom use. The load-equilization performance of the surge tank, obviously, leaves much to be desired.

Sludge Characteristics

For the 3 days of the aeration train profiles, the SVI averaged approximately 100 mℓ/g. ML volatility was about 65 percent. With other conditions remaining unchanged, percent volatility will mirror the iron content of sludge, decreasing as the iron concentration increases.

Oxygen Uptake Rates (OURs)

In conjunction with the profile of December 5, 1973, OURs on the ML were conducted with the following results:

Cell No.	1	2	3	4	5	6	7/8
OUR (mg/O_2/hr/g MLSS)	22.3	9.1	11.1	9.4	7.5	6.5	6.9

Cell No.	9	10	11	12
OUR (mg/O_2/hr/g MLSS)	7.9	7.1	5.4	6.0

Underlined cells were anoxic.

Turbulence in Anoxic Cells

In Cell 7/8, the floc particles were clearly visible and distinguishable from the supernatant. That was not the case with respect to the other anoxic cells. The same condition prevailed in later flow sheets. In Cell 7/8, ML solids were kept in suspension by a mixer. In the other anoxic cells, throttled air was used.

OPERATING RESULTS: FLOW SHEET II

This was the first flow sheet under which the 7.45-kW (10-hp) floating aerator was utilized in the surge tank. The sludge scraper used with Flow Sheet I had been dismantled. The surge tank aerator was operated continuously during the 1 wk Flow Sheet II was evaluated. All other operating aspects of Flow Sheet II were identical to Flow Sheet I (see Figure 13).

Rationale

It had been hoped that the lengthy preaeration period provided by the surge tank would greatly improve the overall performance of the treatment system by:

a) Nearly saturating the influent to the primary settler with DO, thereby alleviating the rising sludge problem in that unit.

b) Effecting a considerable reduction in BOD in the process water, thereby reducing the organic load on the A.S. aeration system and improving its performance. This was supported by literature reports on the effects of preaeration (20).

The improvements hoped for would be brought about without additional tankage.

Results

Utilization of the surge tank for preaeration had a very unfavorable effect and had to be discontinued after a few days. The chlorinated final effluent became "milky," containing a high concentration of SS with correspondingly high BOD_5 and COD concentrations. Some of the analytical results for Flow Sheet II are shown in Table 12.

Full-time aeration of the surge tank was accompanied by heavy foam formation in that treatment unit. The process water in the surge tank and primary settler became more transparent and assumed a brownish tint.

Discussion

It is believed that the "milky" effluent was due to dispersed growth, that full-time aeration in the surge tank, where the average detention time was 9 hr, greatly increased biological action in the surge tank, stimulating bacterial growth there. But the physiological state of the bacterial growth developed therein was such that they did not flocculate nor could they be captured by the A.S. floc.

TABLE 12. OPERATING AND EFFLUENT DATA FOR FLOW SHEET II

	Chlorinated Effluent†				Mixed Liquor		
Date 1974	TSS (mg/ℓ)	BOD_5** (mg/ℓ)	COD (mg/ℓ)	TKN (mg/ℓ)	TSS (mg/ℓ)	SVI (mℓ/g)	Vol. (%)
4/29	7.0	8.5	N.A.	6.3	2240	98	68.8
4/30	24.0	13.7*	65*	9.0*	2790	87	66.2
5/1	40.0*	15.0*	69*	13.2*	2455	99	68.8
5/2	21.0*	18.0*	91*	13.4*	2000	100	70.0
5/3	52.0*	24.0*	80*	8.4*	2380	92	69.6
5/6	110.0	8.5	N.AV.	2.9	3850	117	60.0
5/7	26.0*	28*	83*	5.3*	3085	91	67.0

† Average chlorine dose was 9.5 mg/ℓ.

* Values determined on 24-hr composite samples; values not starred determined as grab samples.

** BOD_5 samples were seeded before analysis.

OPERATING RESULTS: FLOW SHEET III

This flow sheet, depicted schematically in Figure 18, was characterized by the use of the surge tank as part of the activated sludge system and by the absence of primary settling. Raw sewage and return sludge were both introduced to the surge tank, which was operated with the 7.45-kW (10-hp) aerator as an aeration device. The entire compartmentalized aeration tank was also part of the aeration system. The surge tank continued to function as a flow equalization unit.

Rationale

It was felt that using the surge tank as part of the activated sludge system would increase the MLSS mass by approximately 125 percent, thereby drastically lowering the F/M load on the system and improving carbonaceous oxidation and nitrification. Since the surge tank would continue to be used for equalization, this improvement could be brought about without additional capital cost. The additional cost of the floating aerator and larger pumping equipment would be offset by eliminating the cost of the sludge scraper.

Results

The above rationale proved to be correct but not sufficiently comprehensive.

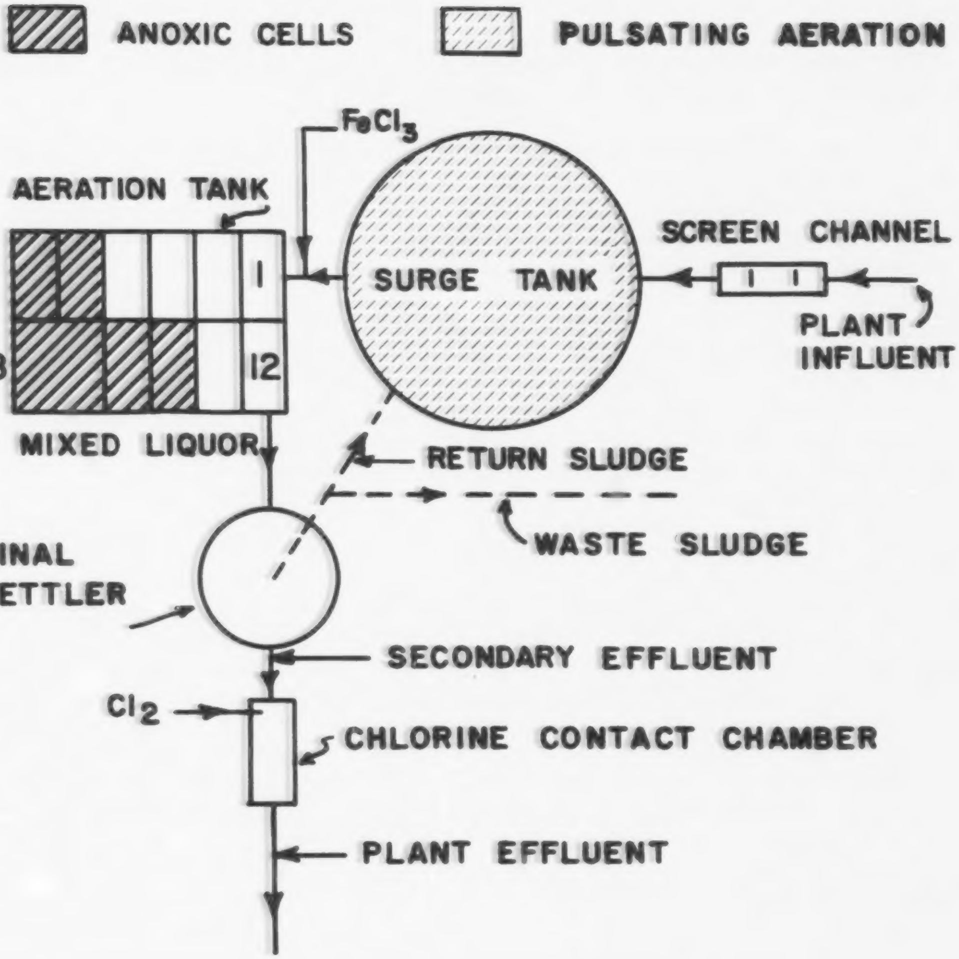

ANOXIC CELLS
PULSATING AERATION
$FeCl_3$
AERATION TANK
6
1
7/8
12
SURGE TANK
SCREEN CHANNEL
PLANT INFLUENT
MIXED LIQUOR
RETURN SLUDGE
WASTE SLUDGE
FINAL SETTLER
SECONDARY EFFLUENT
Cl_2
CHLORINE CONTACT CHAMBER
PLANT EFFLUENT

Figure 18. Process schematic for Flow Sheet III.

Effluent quality at the beginning of this phase was excellent. However, the process soon became unmanageable due to a rising SVI, which reached nearly 500 mℓ/g within 6 wk as indicated in Table 13 . The effect of the new flow sheet on sludge settleability had not been taken into consideration.

Sludge bulking was caused by gram-negative filamentous organisms. A greatly increased population of stalked ciliates was also noticeable. It is believed that the sludge bulking may have been initiated by the switch to a system that included as a first treatment unit a CM reactor of about 9 hr sewage detention time. The combination of a relatively high DO concentration and a low COD concentration in the process water may have triggered a population explosion of the filamentous organisms.

The floating aerator was first operated full time for the first 2 days of Flow Sheet III. Full-time operation of the aerator resulted in a final effluent nitrate concentration of near 10 mg/ℓ N. Under the prevailing operating conditions, this high nitrate nitrogen concentration was accompanied by problems of floating and rising sludge in the final settler. After the first 2 days, the floating aerator was put on a 10 min "on", 10 min "off" cycle. This change in aeration pattern immediately solved the problem of excessive nitrates in the final effluent by allowing substrate nitrate respiration to occur directly in the surge tank. Improved nitrogen removal was accomplished by making use of the wastewater organic carbon. Reduction of the aeration "on" time also obviated a foaming problem in the surge tank. The 10 min "on", 10 min "off" aerator cycle was selected because it was felt that settling of the activated sludge floc would just about be completed in 10 min at low process water levels in the surge tank. The biomass must be in suspension in order for substrate nitrate respiration to occur.

Removals were of course greatly affected by the poor settling qualities of the sludge which frequently resulted in high TSS, BOD_5, and TKN concentrations in the final effluent (see Table 13). Apart from this, the results indicate the probability limits to which nitrogen removal can be pushed in a one-sludge system without a supplemental carbon source. TKN probably cannot be reduced below 1.5 to 2.0 mg/ℓ. $NO_2^- + NO_3^-$-N probably cannot be consistently reduced to below 0.5 mg/ℓ. Such performance will require utilization of a manageable sludge blanket in the final settler, not the bulking high sludge blanket experienced in this phase.

Conclusions

The utilization of a surge tank as part of an aerobic-anoxic A.S. system appears to be a viable arrangement, particularly for smaller plants. The surge tank provides both flow equalization, which is important for effective final settler operation, and substrate nitrate respiration, which is important for effective nitrogen removal. Provisions for coping with sludge bulking must be made, however.

The floating aerator should be activated by a time clock and have a direct drive. Gear-driven aerators are not suited for the "on-off" operation required.

TABLE 13. OPERATING AND EFFLUENT DATA FOR FLOW SHEET III

Date 1974	Flow (m^3/day)	Mixed Liquor MLSS (mg/ℓ)	Mixed Liquor SVI (mℓ/g)	Effluent, Chlorine Contact Chamber† TSS (mg/ℓ)	BOD_5** (mg/ℓ)	TKN (mg/ℓ)	NO_2+NO_3-N (mg/ℓ)
5/8	416	2040	83	2.0	8.0	N.AV.	10.9
5/9	454	2140	77	8.0*	15.0	3.4*	8.5*
5/10	416	2675	83	12.4	25.0*	2.2*	2.4*
5/13	568	2090	86	16.0	8.5	2.5	5.0
5/14	378	2840	92	21.5*	13.7	2.9*	3.9*
5/15	341	3750	186	23.0*	16.0*	2.5*	2.8*
5/16	341	2990	193	13.0*	N.AV.	2.1*	0.7*
5/17	341	1695	100	11.5*	12.0*	2.0*	1.8*
5/20	341	985	102	6.2	N.AV.	1.7	3.6
5/21	N.AV.	3900	210	314*	N.AV.	22.9*	1.4*
5/22	530	2890	250	69*	N.AV.	3.6*	0.01*
5/23	530	3390	208	9*	N.AV.	1.1*	0.01*
5/24	492	4000	68	75*	N.AV.	1.7*	0.2*
5/27	492	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.
5/28	454	4385	220	3.0	3.4	0.8	0.2
5/29	530	4260	370	6.5	14*	0.7	0.1
5/30	530	3520	320	7.0*	7*	1.4*	0.4*
5/31	568	2705	245	8.5*	10*	1.5*	0.4*
6/3	568	4150	234	14.0	6.0	1.2	0.1
6/4	530	3410	276	1.5*	10.3*	2.2*	0.2*
6/5	568	3995	242	7.2*	15.0*	N.AV.	0.2*
6/6	568	1450	207	14.5*	12.0*	2.1*	0.6*
6/7	530	2840	N.AV.	6.0*	8.0*	1.1	0.5
6/10	530	3200	294	3.0*	6.9	0.8	0.4
6/11	606	2940	310	44.0*	36.0*	3.78*	0.2*
6/12	568	2960	318	9.0*	12.0*	1.68*	0.7*
6/13	568	3060	307	4.2*	5.0*	1.7*	0.7*
6/14	568	3270	291	24.0*	15.0*	2.1*	0.4*
6/17	606	3005	N.AV.	6.0	10.0	1.8	0.3
6/18	530	2235	425	33.0*	22.0*	2.2*	0.3*
6/19	606	2405	401	26.0*	23.0*	4.3*	0.2*
6/20	530	2395	401	36.0*	25.0*	2.4*	0.2*
6/21	568	1945	478	195.0*	N.AV.	14.6*	0.2*

† Average chlorine dose was 9.0 mg/ℓ.

* Values determined on 24-hr composite samples; values not starred determined on grab samples.

** BOD_5 samples were seeded before analysis.

Average detention time in surge tank: approximately 9 hr.

The surge tank pumps should be suitable for pumping ML. They have to accommodate both the plant flow as well as the return sludge flow.

TERMINOLOGY USED IN TABLES DESCRIBING PERFORMANCE OF FLOW SHEETS IV, V, AND VI

The bulk of data describing the operating conditions and performance results for Flow Sheets IV, V, and VI (Figures 19, 20, and 21, respectively) are contained in Tables 14 through 31 and Figure 22. The following remarks explain how some of these data were computed.

"Daily Flow" is the daily flow introduced into the primary settler or the aeration tank via the variable sluice gate in the sideweir structure, averaged over one calendar month. The daily flows for the month were added and the sum divided by the number of days in the month.

"Mean Sewage Detention Time," for all units except the surge tank, is the average monthly nominal fluid retention time. It was found by dividing the "Daily Flow", as described above, into the volumetric capacity of the treatment unit. "Monthly High (Low) Sewage Detention Times" were determined by dividing the highest (lowest) daily plant flow of the calendar month into the capacity of the treatment unit.

Similar procedures were followed when determining the other hydraulic parameters based on plant flow: surface overflow rate and weir overflow rate.

"Mean Sewage Detention Time" for the surge tank was determined as the flow weighted monthly average of the daily sewage detention time in the surge tank using the same method of computation utilized for determining the monthly flow weighted average of substrate concentrations found in the process water. "High" and "Low" sewage detention times for the month were found simply by inspecting the list of daily detention times for the calendar month under consideration. Determination of the surge tank daily sewage detention time is described in Section 6.

"Mixed Liquor Detention Time" was computed by dividing the average monthly ML flow into the capacity of the aeration tank. ML flow is the sum of sewage flow and sludge recycle flow.

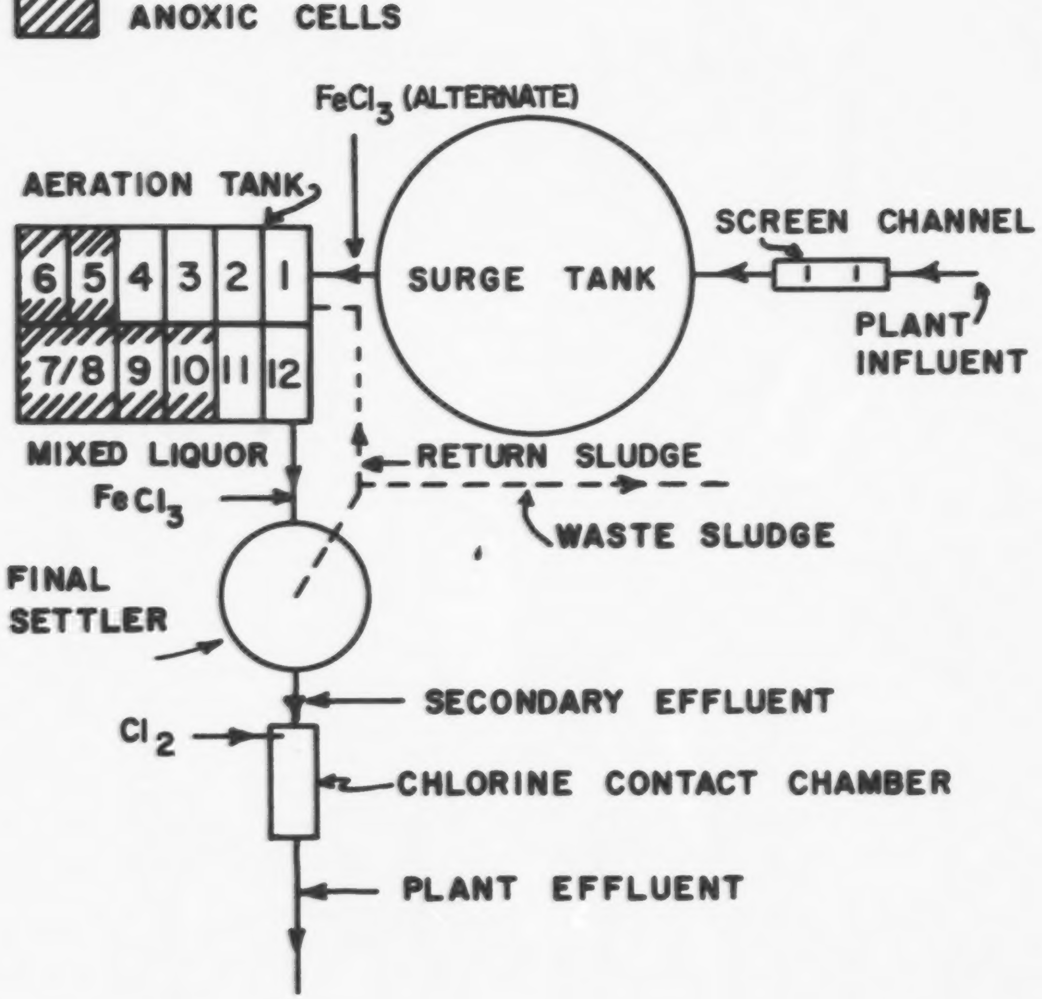
Monthly flow weighted average concentrations C_m were determined by using the equation:

$$C_m = \frac{\Sigma C_n Q_n}{\Sigma Q_n} \qquad (38)$$

where: C_n = concentration, 24-hr composite, n^{th} day of the month, mg/ℓ

Q_n = daily flow, n^{th} day of the month

Σ = summation for the days of the month.



▨ ANOXIC CELLS
FeCl3 (ALTERNATE)
AERATION TANK
6
5
4
3
2
1
7/8
9
10
11
12
SURGE TANK
SCREEN CHANNEL
PLANT INFLUENT
MIXED LIQUOR
FeCl3
RETURN SLUDGE
WASTE SLUDGE
FINAL SETTLER
SECONDARY EFFLUENT
Cl2
CHLORINE CONTACT CHAMBER
PLANT EFFLUENT

Figure 19. Process schematic for Flow Sheet IV.

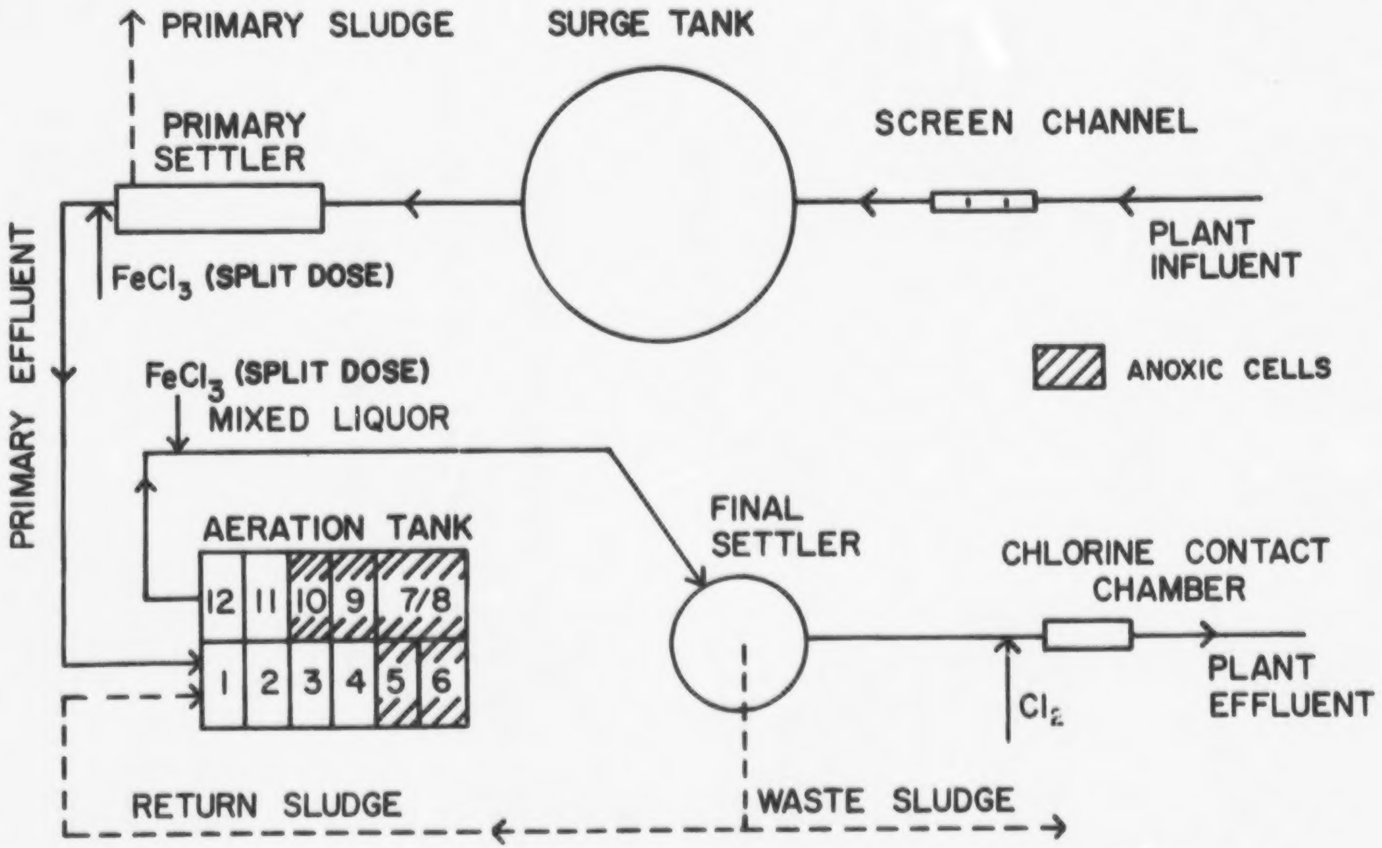
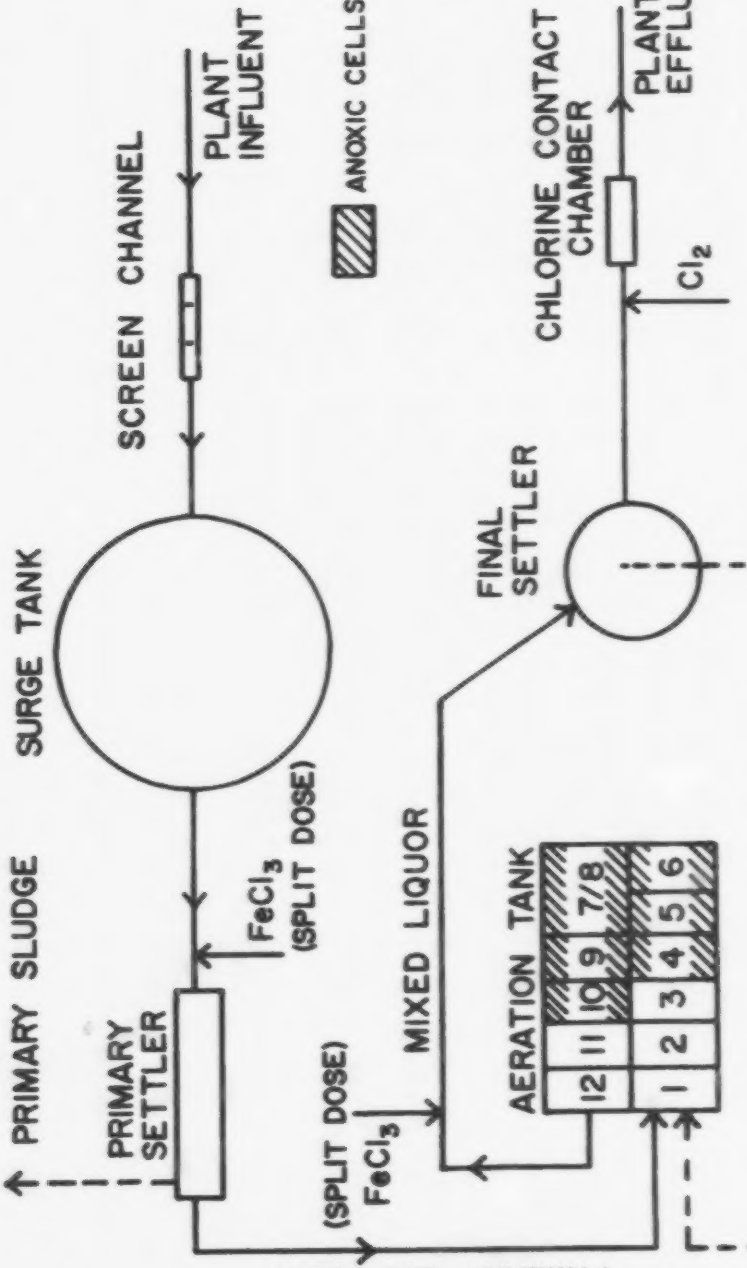

PRIMARY SLUDGE
SURGE TANK
PRIMARY SETTLER
SCREEN CHANNEL
PLANT INFLUENT
FeCl$_3$ (SPLIT DOSE)
PRIMARY EFFLUENT
ANOXIC CELLS
FeCl$_3$ (SPLIT DOSE)
MIXED LIQUOR
AERATION TANK
12
11
10
9
7/8
1
2
3
4
5
6
FINAL SETTLER
CHLORINE CONTACT CHAMBER
Cl$_2$
PLANT EFFLUENT
RETURN SLUDGE
WASTE SLUDGE

Figure 20. Process schematic for Flow Sheet V.



PRIMARY SLUDGE
PRIMARY SETTLER
SURGE TANK
SCREEN CHANNEL
PLANT INFLUENT
FeCl3 (SPLIT DOSE)
ANOXIC CELLS
MIXED LIQUOR
(SPLIT DOSE) FeCl3
AERATION TANK
12
11
10
9
7/8
1
2
3
4
5
6
FINAL SETTLER
CHLORINE CONTACT CHAMBER
Cl2
PLANT EFFLUENT
RETURN SLUDGE
WASTE SLUDGE
PRIMARY EFFLUENT

Figure 21. Process schematic for Flow Sheet VI.

TABLE 14. AIR FLOW TO PLANT AND DO IN AERATION CELLS DURING FLOW SHEET IV, V, AND VI OPERATIONS

Year	Month	Flow Sheet	Air Flow to Aeration Cells (ℓ/sec)	Air Flow to Aeration Cells (scfm)	DO in Aeration Cells (mg/ℓ) Cell 1	Cell 2	Cell 3	Cell 4	Cell 5	Cell 6	Cell 7/8	Cell 9	Cell 10	Cell 11	Cell 12	Sewage Flow (m^3/day)
1974	Aug.	IV	176	(372)	3.7	4.0	5.0	5.0	0.4	0.1	0.1	0.1	1.3	3.6	5.1	541
1974	Sept.	IV	86	(182)	1.8	2.9	3.2	3.7	0.2	0.2	0.2	0.1	0.1	3.7	4.5	598
1974	Oct.	IV	112	(237)	2.9	3.7	3.9	4.8	0.3	0.2	0.2	0.2	0.2	5.2	5.2	462
1974	Nov.	IV	125	(265)	3.6	4.3	4.6	5.2	0.4	0.4	0.2	0.6	0.6	5.2	7.0	534
1974	Dec.	IV	114	(241)	1.5	3.8	4.5	4.7	0.4	0.2	0.1	0.3	0.3	6.3	6.9	488
1975	Jan.	IV	140	(296)	4.2	4.5	4.6	5.2	0.5	0.2	0.3	0.3	0.3	4.5	5.9	492
1975	Feb.	IV	132	(280)	4.2	4.6	4.4	4.6	0.5	0.2	0.2	0.2	0.2	6.5	6.0	454
1975	Mar.	IV	115	(243)	4.5	4.0	3.9	4.0	0.3	0.2	0.1	0.1	0.1	5.2	5.9	477
1975	Apr.	V	131	(278)	4.2	4.4	4.2	4.2	1.1	0.3	0.2	0.2	0.3	3.2	4.8	488
1975	May	VI	142	(300)	3.5	3.7	3.8	2.8	0.5	0.2	0.2	0.2	0.2	0.9	6.5	492
1975	June	VI	116	(246)	4.5	4.0	3.9	4.0	0.3	0.2	0.1	0.1	0.1	5.2	5.9	590
1975	July	VI	117	(248)	2.7	0.1	2.0	0.1	1.8	0.1	0.1	2.5	0.2	0.1	6.1	643

TABLE 15. MONTHLY HYDRAULIC SURGE TANK, PRIMARY SETTLER, AND CHLORINE CONTACT CHAMBER DATA FOR FLOW SHEETS IV, V, AND VI

			Surge Tank			Primary Settler					C.C.C.**
			Sewage Detention Time			S.D.T.*	Surface Overflow Rate		Weir Overflow Rate		S.D.T.*
Month	Year	Flow Sheet	Mean (hr)	High (hr)	Low (hr)	Mean (hr)	Mean (m^3/m^2/day)	Mean (gpd/ft^2)	Mean (m^3/m/day)	Mean (gpd/ft)	Mean (hr)
Aug.	1974	IV	8.2	10.9	5.5	N.AP.	N.AP.	N.AP.	N.AP.	N.AP.	0.7
Sept.	1974	IV	8.2	11.7	5.6	N.AP.	N.AP.	N.AP.	N.AP.	N.AP.	0.6
Oct.	1974	IV	8.4	12.1	3.3	N.AP.	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Nov.	1974	IV	10.5	14.5	8.0	N.AP.	N.AP.	N.AP.	N.AP.	N.AP.	0.7
Dec.	1974	IV	12.0	16.3	8.4	N.AP.	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Jan.	1975	IV	11.2	21.4	6.6	N.AP.	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Feb.	1975	IV	12.2	18.9	6.7	N.AP.	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Mar.	1975	IV	11.9	20.7	4.2	N.AP.	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Apr.	1975	V	11.1	15.8	6.1	1.7	35	(870)	260	(21,000)	0.8
May	1975	VI	11.8	16.7	9.0	1.7	36	(880)	260	(21,000)	0.8
June	1975	VI	11.1	16.8	6.9	1.4	43	(1050)	310	(25,000)	0.6
July	1975	VI	8.6	12.8	5.9	1.2	47	(1160)	350	(28,000)	0.6

* S.D.T. = Sewage Detention Time.

** C.C.C. = Chlorine Contact Chamber.

TABLE 16. MONTHLY-AVERAGE AERATION TANK AND FINAL SETTLER DETENTION TIMES FOR FLOW SHEETS IV, V, AND VI

			Aeration Tank						Final Settler		
			Sewage Detention Time			Mixed Liquor Detention Time			Sewage Detention Time		
Month	Year	Flow Sheet	Mean (hr)	Low (hr)	High (hr)	Mean (hr)	Low (hr)	High (hr)	Mean (hr)	Low (hr)	High (hr)
Aug.	1974	IV	6.6	5.4	8.2	4.9	4.2	5.8	2.6	2.1	3.2
Sept.	1974	IV	6.0	4.5	8.3	4.6	3.6	5.7	2.3	1.8	3.2
Oct.	1974	IV	7.8	5.5	9.5	5.7	4.4	6.6	3.0	2.3	3.7
Nov.	1974	IV	6.8	4.5	8.9	5.2	3.5	6.6	2.7	1.8	3.5
Dec.	1974	IV	7.3	4.9	9.3	5.3	3.9	6.5	2.8	2.1	3.6
Jan.	1975	IV	7.5	5.2	11.4	5.3	3.9	7.6	2.9	2.1	4.5
Feb.	1975	IV	8.1	5.3	14.2	5.7	4.1	8.6	3.2	2.3	5.5
Mar.	1975	IV	7.6	5.4	12.0	5.4	4.0	7.9	3.0	2.3	4.7
Apr.	1975	V	7.4	5.3	9.2	5.6	4.2	6.7	2.9	2.1	3.6
May	1975	VI	7.3	5.2	9.6	5.5	3.9	7.1	2.9	2.0	3.7
June	1975	VI	6.1	4.8	9.5	4.7	3.8	6.8	2.4	1.9	3.7
July	1975	VI	5.5	4.8	7.4	4.3	3.7	5.6	2.1	1.9	2.9

TABLE 17. MONTHLY-AVERAGE FINAL SETTLER OVERFLOW AND SOLIDS LOADING RATES FOR FLOW SHEETS IV, V, AND VI

			Surface Overflow Rate			Surface Overflow Rate			Weir Overflow Rate		Solids Loading Rate*	
			(m^3/m^2/day)			(gpd/ft^2)			(m^3/m/day)	(gpd/ft)	(kg/m^2/day)	(lb/ft^2/day)
Month	Year	Flow Sheet	Mean	High	Low	Mean	High	Low	Mean	Mean	Mean	Mean
Aug.	1974	IV	29	35	23	(711)	(850)	(565)	41	(3300)	113	(23)
Sept.	1974	IV	31	42	23	(760)	(1030)	(555)	46	(3700)	113	(23)
Oct.	1974	IV	24	34	20	(588)	(835)	(485)	35	(2800)	118	(24)
Nov.	1974	IV	27	42	21	(662)	(1020)	(520)	40	(3200)	100	(20)
Dec.	1974	IV	26	38	20	(638)	(940)	(495)	37	(3000)	117	(24)
Jan.	1975	IV	25	36	17	(613)	(890)	(405)	36	(2900)	129	(26)
Feb.	1975	IV	23	35	13	(564)	(865)	(325)	34	(2700)	105	(21)
Mar.	1975	IV	25	35	16	(613)	(860)	(385)	36	(2900)	125	(26)
Apr.	1975	V	25	35	20	(613)	(865)	(500)	37	(3000)	120	(25)
May	1975	VI	26	36	20	(638)	(890)	(480)	37	(3000)	125	(26)
June	1975	VI	31	39	20	(760)	(960)	(485)	45	(3600)	143	(29)
July	1975	VI	34	39	25	(834)	(965)	(620)	49	(3900)	167	(34)

$$\text{* Solids Load Rate} = \frac{\text{MLSS (mg/}\ell\text{) x Mixed Liquor Flow (m}^3\text{/day)}}{\text{Final Settler Area (m}^2\text{) x }10^3}\ ;\ \text{kg/m}^2\text{/day}$$

$$= \frac{\text{8.34 MLSS (mg/}\ell\text{) x Mixed Liquor Flow (mgd)}}{\text{Final Settler Area (ft}^2\text{)}}\ ;\ \text{lb/ft}^2\text{/day}$$

TABLE 18. FLOW-WEIGHTED, MONTHLY-AVERAGE SURGE TANK EFFLUENT CHARACTERISTICS FOR FLOW SHEETS IV, V, AND VI

Parameter	Aug. 1974	Sept. 1974	Oct. 1974	Nov. 1974	Dec. 1974	Jan. 1975	Feb. 1975	Mar. 1975	Apr. 1975	May 1975	June 1975	July 1975
Flow Sheet	IV	IV	IV	IV	IV	IV	IV	IV	V	VI	VI	VI
Daily Flow (m^3)	534	583	451	511	477	466	432	458	473	477	572	633
Daily Flow (mil gal)	(0.141)	(0.154)	(0.119)	(0.135)	(0.126)	(0.123)	(0.114)	(0.121)	(0.125)	(0.126)	(0.151)	(0.167)
Temperature (C)	26.0	24.1	23.8	20.7	18.0	18.8	17.8	20.5	19.7	23.1	24.8	27.6
pH	7.2	7.2	7.2	7.4	7.4	7.4	7.5	7.4	7.4	7.4	7.5	7.4
Alk. (mg/ℓ as $CaCO_3$)	168	179	190	204	219	212	229	206	197	212	217	172
SS (mg/ℓ)	205	241	270	223	276	241	242	267	250	246	253	218
COD (mg/ℓ)	413	378	433	398	396	434	457	464	457	434	429	386
BOD_5 (mg/ℓ)	181	163	195	186	194	208	225	213	189	177	157	142
Diss. BOD_5 (mg/ℓ)	63	59	66	53	N.AV.	92	75	75	73	60	47	40
NH_4^+-N (mg/ℓ)	10.5	9.6	10.9	12.3	12.0	12.8	14.9	14.7	13.7	16.3	15.1	11.0
Organically Bound N (mg/ℓ)	15.6	12.3	14.8	12.1	14.5	15.6	17.1	15.5	15.4	13.9	13.9	12.3
TKN (mg/ℓ)	26.0	22.3	26.1	24.4	26.2	28.4	32.0	30.2	29.3	30.3	29.0	23.3
NH_4^+-N as % of TKN	41	43	42	50	46	45	47	49	47	54	52	47
NO_2^-+NO_3^--N (mg/ℓ)	0.25	0.2	0.01	0.05	0.14	0.07	0.13	0.13	0.25	0.31	0.24	0.15
Total N (mg/ℓ)	26.0	22.5	26.1	24.5	26.3	28.7	32.1	30.4	29.5	30.6	29.2	23.5
TKN as % of COD	6	6	6	6	7	7	7	7	6	7	7	6
Ratio COD/TKN	16	17	17	16	15	15	14	15	16	14	15	17
Diss. Orthophosphorus (mg/ℓ)	6.4	4.3	4.0	4.5	4.9	4.6	5.3	5.5	5.5	4.5	4.4	4.0
Diss. Phosphorus (mg/ℓ)	7.1	4.6	4.3	4.4	4.9	4.6	5.0	5.5	4.1	N.AV.	N.AV.	N.AV.
Total Phosphorus (mg/ℓ)	11.4	9.7	9.9	9.1	10.3	9.6	10.2	10.8	10.0	9.2	9.1	8.3
DO (mg/ℓ)	N.AV.	0.3	0.4	0.4	0.4	0.3	0.4	0.4	0.3	0.3	0.2	0.3

TABLE 19. MONTHLY-AVERAGE AERATION TANK LOADING RATES AND SOLIDS RETENTION TIMES FOR FLOW SHEETS IV, V, AND VI

Month	Year	Flow Sheet	Solids Retention(1) Time Based on		Aerobic Solids(2) Retention Time	F/M Load(3) Based on		Volumetric Organic Load(3)	
			MLSS (days)	MLVSS (days)	Based on MLSS (days)	MLVSS (kg/day/kg)	MLSS (kg/day/kg)	(kg/day/m^3)	(lb/day/1000 ft^3)
Aug.	1974	IV	4.4	4.4	2.2	0.35	0.23	0.67	42
Sept.	1974	IV	3.7	3.6	1.9	0.37	0.23	0.66	41
Oct.	1974	IV	5.5	5.5	2.8	0.26	0.17	0.59	37
Nov.	1974	IV	3.8	3.9	1.9	0.34	0.23	0.66	41
Dec.#	1974	IV	4.8#	5.4#	2.4#	0.28#	0.19#	0.64#	40#
Jan.	1975	IV	6.1	6.1	3.0	0.26	0.18	0.66	41
Feb.	1975	IV	4.5	4.6	2.3	0.28	0.20	0.67	42
Mar.	1975	IV	5.2	5.4	2.6	0.27	0.18	0.66	41
Apr.	1975	V	6.2	6.1	3.1	0.23	0.14	0.50	31
May	1975	VI	8.1	8.2	3.4	0.17	0.11	0.42	26
June	1975	VI	7.0	6.9	2.9	0.19	0.12	0.43	27
July	1975	VI	9.0	8.7	3.8	0.17	0.10	0.38	24

December data are of little significance because plant was operated as a primary plant for part of the month due to breakdown of final settler equipment.

(1) Based on solids in entire aerator.

(2) Based on BOD_5 in influent to aeration tank and solids in aerobic cells only.

(3) Based on BOD_5 in influent to aeration tank and on entire aeration tank volume.

TABLE 20. MONTHLY-AVERAGE MIXED LIQUOR CHARACTERISTICS FOR FLOW SHEETS IV, V, AND VI

Month	Year	Flow Sheet	Sludge Recycle Ratio (%)	MLSS (mg/ℓ)	MLVSS (mg/ℓ)	% Volatile	SVI (ml/g)	*SV60/ SV30	Fe^{+++} (% of MLSS)	TKN (% of MLSS)	Fe^{+++} (% of MLVSS)	TKN (% of MLVSS)
Aug.	1974	IV	34	2950	1900	0.65	108	0.77	6.9	6.0	11.0	9.0
Sept.	1974	IV	30	2800	1770	0.63	91	0.78	8.1	5.8	13.0	9.3
Oct.	1974	IV	36	3590	2320	0.65	93	0.75	5.2	6.1	8.0	9.4
Nov.	1974	IV	30	2800	1920	0.69	76	0.82	4.2	6.2	6.0	9.0
Dec.	1974	IV	38	3320	2320	0.69	64	0.90	4.3	6.3	6.2	8.5
Jan.	1975	IV	41	3680	2610	0.71	64	0.88	5.6	6.8	7.9	9.5
Feb.	1975	IV	42	3210	2350	0.73	75	0.81	4.1	7.1	5.6	9.7
Mar.	1975	IV	41	3610	2440	0.68	110	0.80	7.0	5.8	10.0	8.6
Apr.	1975	V	33	3560	2210	0.62	73	0.85	8.8	5.4	14.3	10.5
May	1975	VI	33	3690	2390	0.65	88	0.80	7.3	N.AV.	11.1	N.AV.
June	1975	VI	29	3620	2230	0.62	68	N.AV.	9.2	N.AV.	14.8	N.AV.
July	1975	VI	28	3850	2280	0.59	59	N.AV.	10.1	N.AV.	17.1	N.AV.

* SV60/SV30 = Settled sludge volume in 60 min/settled sludge volume in 30 min.

TABLE 21. MONTHLY-AVERAGE EXCESS ACTIVATED SLUDGE CHARACTERISTICS FOR FLOW SHEETS IV, V, AND VI

	Flow Sheet	RSSS (mg/ℓ)	RSVSS (mg/ℓ)	Fe^{+++} (mg/ℓ)	Fe^{+++} (% of RSSS)	TKN (mg/ℓ)	TKN (% of RSSS)	P* (mg/ℓ)	P (% of RSSS)	Molar Ratio Fe/P	RSSS/ MLSS	% Vol.	%# Vol. Adj.
1974													
Aug.	IV	11,400	7200	940	8.6	880	5.3	559	3.9	0.93	3.84	63	75
Sept.	IV	11,100	6900	930	8.4	640	5.8	524	4.7	0.98	3.96	62	74
Oct.	IV	13,300	8500	710	5.4	770	5.8	671	5.1	0.59	3.71	64	71
Nov.	IV	12,000	8100	520	4.3	710	5.9	469	3.9	0.62	4.29	68	74
Dec.	IV	13,300	8500	550	4.1	750	5.6	511	3.8	0.59	4.00	64	69
1975													
Jan.	IV	13,100	9200	760	5.8	840	6.4	475	3.6	0.88	3.56	70	79
Feb.	IV	11,300	8300	420	3.7	750	6.7	368	3.3	0.64	3.52	73	79
Mar.	IV	12,700	8600	810	6.7	680	5.7	537	4.5	0.81	3.52	67	77
Apr.	V	15,000	9500	1320	8.8	810	5.4	620	4.1	1.18	4.21	63	76
May	VI	14,200	11,400	1050	7.4	850	6.0	600	4.2	0.95	3.85	65**	76
June	VI	14,500	9100	1290	8.9	780	5.4	713	4.9	1.00	4.00	63	76
July	VI	16,600	10,300	1690	10.1	890	5.4	833	5.0	1.12	4.31	62	77

* Determined by computation.

** Determined from ML data.

# Before computing % Vol., weight of Fe^{+++} as $Fe(OH)_3$ was deducted from RSSS.

TABLE 22. MONTHLY-AVERAGE SOLIDS PRODUCTION DATA FOR FLOW SHEETS IV, V, AND VI

Month	Year	Flow Sheet	EAS (liter per liter of sewage)	EAS (mg of SS per liter of sewage)	EAS (mg of VSS per liter of sewage)	EAS (mg of SS per mg of BOD_5 removed)	EAS (mg of VSS per mg of BOD_5 removed)	Solids* (mg of SS per liter of sewage treated)	Solids* (mg of VSS per liter of sewage treated)	Solids* (mg of SS per mg of BOD_5 removed)	Solids* (mg of VSS per mg of BOD_5 removed)
Aug.	1974	IV	0.017	180	114	1.01	0.64	186	118	1.04	0.66
Sept.	1974	IV	0.017	176	114	1.14	0.73	189	122	1.22	0.79
Oct.	1974	IV	0.016	203	131	1.07	0.68	211	136	1.11	0.72
Nov.	1974	IV	0.017	199	133	1.12	0.75	209	140	1.18	0.79
Dec.#	1974	IV	0.009#	143#	88#	0.9#	0.54#	207#	130#	1.3#	0.79#
Jan.	1975	IV	0.017	208	146	1.07	0.75	228	161	1.18	0.83
Feb.	1975	IV	0.020	222	158	1.09	0.76	242	173	1.19	0.85
Mar.	1975	IV	0.017	202	135	1.05	0.70	215	144	1.12	0.75
Apr.	1975	V	0.011	165	104	1.15	0.72	176	111	1.23	0.78
May	1975	VI	0.009	126	81	1.10	0.70	138	89	1.20	0.77
June	1975	VI	0.009	125	78	1.23	0.77	132	82	1.29	0.80
July	1975	VI	0.006	93	57	1.08	0.66	98	60	1.12	0.70

* "Solids" include EAS solids and solids in final effluent.

# December data are of little significance because plant was operated as a primary plant for part of the month due to breakdown of final settler equipment.

TABLE 23. FLOW-WEIGHTED, MONTHLY-AVERAGE FINAL SETTLER EFFLUENT CHARACTERISTICS FOR FLOW SHEETS IV, V, AND VI

Parameter	Aug. 1974	Sept. 1974	Oct. 1974	Nov. 1974	Dec. 1974	Jan. 1975	Feb. 1975	Mar. 1975	April 1975	May 1975	June 1975	July 1975
Flow Sheet	IV	IV	IV	IV	IV	IV	IV	IV	V	VI	VI	VI
Daily Flow, Surge Tank Eff. (m^3)	534	583	451	511	477	466	432	458	473	477	572	633
Daily Flow, Surge Tank Eff. (mil gal)	(0.141)	(0.154)	(0.119)	(0.135)	(0.126)	(0.123)	(0.114)	(0.121)	(0.125)	(0.126)	(0.151)	(0.167)
Temperature of Fin. Set. Eff. (C)	25.5	22.5	21.5	19.3	16.3#	16.6	15.9	18.1	18.5	23.0	24.5	27.1
SS (mg/ℓ)	6	13	8	10	64#	21	20	13	11	12	7	5
VSS (mg/ℓ)	N.AV.	9	5	6	15#	11	16	10	7	8	4	3
% Vol. SS	N.AV.	69	63	60	23#	52	80	77	64	67	57	60
COD (mg/ℓ)	35	47	38	52	101#	48	61	60	66	61	54	51
Diss. COD (mg/ℓ)	31	30	33	31	N.AV.	37	37	46	52	50	50	46
BOD_5 (mg/ℓ)	6	8	5	9	30#	14	21	18	12	11	8	2
Diss. BOD_5 (mg/ℓ)	3	3	3	2	N.AV.	5	4	7	4	4	1	2
NH_4^+-N (mg/ℓ)	0.4	0.6	0.3	2.0	3.2#	0.8	1.2	2.7	2.1	0.5	0.2	0.1
Organically Bound N (mg/ℓ)	1.3	1.4	1.7	2.4	5.3#	2.7	3.3	2.3	2.0	2.2	1.7	1.4
TKN (mg/ℓ)	1.7	2.4	2.0	4.5	8.5#	3.4	4.5	4.8	4.1	2.7	1.9	1.5
NO_2^-+NO_3^--N (mg/ℓ)	1.9	1.4	1.4	1.3	2.4#	1.3	1.0	1.7	3.8	3.8	4.4	3.6
Total N (mg/ℓ)	3.6	3.8	3.3	5.8	10.9#	4.7	5.4	6.5	7.9	6.5	6.3	5.2
Iron (mg/ℓ)	0.1	0.6	1.4	0.1	1.6#	1.4	0.2	0.7	1.4	1.5	1.7	1.8
Diss. Orthophosphorus (mg/ℓ)	1.8	0.4	0.2	0.7	2.1#	1.0	2.3	1.2	1.9	1.9	1.0	0.6
Diss. Phosphorus (mg/ℓ)	1.7	0.4	0.2	0.6	2.0#	1.0	2.1	1.1	2.1	1.6	1.2	0.7
Total Phosphorus (mg/ℓ)	2.0	0.8	0.5	1.6	4.6#	2.0	3.2	1.9	2.7	2.4	1.5	0.9
pH	7.5	7.4	7.4	7.4	7.5#	7.4	7.5	7.5	7.4	7.5	7.4	7.3
Alk. (mg/ℓ as $CaCO_3$)	93	116	133	145	161#	135	151	119	102	111	107	79
DO (mg/ℓ)	N.AV.	1.1	1.5	1.7	1.5#	1.2	0.8	0.8	1.2	1.3	2.2	1.6

December data are of little significance because plant was operated as a primary plant for part of the month due to breakdown of final settler equipment.

TABLE 24. MONTHLY-AVERAGE COD AND BOD_5 REMOVAL EFFICIENCIES FOR FLOW SHEETS IV, V, AND VI

			COD Removal					BOD_5 Removal				
Month	Year	Flow Sheet	Surge Tank (%)	Primary Settler (%)	Aeration System (%)	Chl. Cont. Chamber (%)	Total Plant (%)	Surge Tank (%)	Primary Settler (%)	Aeration System (%)	Chl. Cont. Chamber (%)	Total Plant (%)
Aug.	1974	IV	14	N.AP.	92	N.AP.	N.AP.	17	N.AP.	97	N.AP.	N.AP.
Sept.	1974	IV	10	N.AP.	88	30	92	14	N.AP.	95	62	98
Oct.	1974	IV	4	N.AP.	91	13	93	10	N.AP.	97	40	99
Nov.	1974	IV	14	N.AP.	87	15	90	22	N.AP.	95	56	98
Dec.	1974	IV	30	N.AP.	74	5	83	22	N.AP.	84	13	90
Jan.	1975	IV	35	N.AP.	89	4	93	35	N.AP.	93	36	97
Feb.	1975	IV	23	N.AP.	87	8	91	26	N.AP.	91	43	96
Mar.	1975	IV	16	N.AP.	87	13	90	22	N.AP.	91	44	96
Apr.	1975	V	27	22	81	8	90	32	18	92	58	98
May	1975	VI	18	26	81	10	90	27	28	91	36	97
June	1975	VI	21	31	82	7	91	26	30	93	75	99
July	1975	VI	14	36	79	10	90	17	38	98	0	99

For treatment units, efficiency is based on unit influent contaminant concentrations; for entire plant, efficiency is based on raw sewage contaminant concentrations.

TABLE 25. MONTHLY TREATMENT SYSTEM NITROGEN BALANCES (EXCLUDING SURGE TANK AND CHLORINE CONTACT CHAMBER) FOR FLOW SHEETS IV, V, AND VI

		N-Balance in terms of mg of N applicable to 1 liter of average monthly flow					N-Balance in terms of percent N introduced into plant by surge tank effluent				
	Flow Sheet	Contained in Surge Tank Effluent (mg/ℓ)	Removed in Primary Settler (Sludge) (mg/ℓ)	Denitrified* (mg/ℓ)	Removed in Final Settler (Sludge) (mg/ℓ)	Contained in Secondary Effluent (mg/ℓ)	Contained in Surge Tank Effluent (%)	Removed in Primary Settler (Sludge) (%)	Denitrified* (%)	Removed in Secondary Settler (Sludge) (%)	Contained in Secondary Effluent (%)
1974											
Aug.	IV	26.2	N.AP.	13.1	9.5	3.6	100	N.AP.	50	36	14
Sept.	IV	22.5	N.AP.	8.5	10.2	3.8	100	N.AP.	38	45	17
Oct.	IV	26.1	N.AP.	11.0	11.8	3.3	100	N.AP.	42	45	13
Nov.	IV	24.5	N.AP.	7.0	11.7	5.8	100	N.AP.	28	48	24
Dec.	IV	26.3#	N.AP.	7.4#	8#	10.9#	100#	N.AP.	28#	30#	41#
1975											
Jan.	IV	28.7	N.AP.	10.7	13.3	4.7	100	N.AP.	37.3	46.3	16.4
Feb.	IV	32.1	N.AP.	11.8	14.9	5.4	100	N.AP.	37	46	17
Mar.	IV	29.9	N.AP.	11.9	11.5	6.5	100	N.AP.	40	38	22
Apr.	V	29.5	2.1	10.6	8.9	7.9	100	7	36	30	27
May	VI	30.4	3.1	13.2	7.6	6.5	100	10.2	43.4	25	21.4
June	VI	29.2	3.8	12.4	6.7	6.3	100	13	42	23	22
July	VI	23.5	2.6	10.7	5.0	5.2	100	11	46	21	22

December data are of little significance because plant was operated as a primary plant for part of the month due to breakdown of final settler equipment.

* Determined by deduction. Refers to N removed in ENR.

TABLE 26. MONTHLY TREATMENT SYSTEM Fe^{+++} BALANCES (EXCLUDING CHLORINE CONTACT CHAMBER) FOR FLOW SHEETS IV, V, AND VI

Month	Year	Flow Sheet	Fe^{+++} Contained in Raw Sewage (mg/ℓ) Col. I	Fe^{+++} Introduced Into Treatment Train as $FeCl_3$ (mg/ℓ) Col. II	Fe^{+++} Removed in Primary & Secondary Settler* (mg/ℓ) Col. III	Fe^{+++} Removed in EAS and Primary Sludge: Computed from Tables 21 & 22 (mg/ℓ) Col. IIIa	Fe^{+++} Contained in Secondary Effluent (mg/ℓ) Col. IV	% of Fe^{+++} Recovered in Sludge** Col. V
Aug.	1974	IV	0.3	12.5	12.7	15.5	0.1	122
Sept.	1974	IV	0.3	13.7	13.4	14.8	0.6	110
Oct.	1974	IV	1.5	8.1	8.2	11.0	1.4	129
Nov.	1974	IV	0.3	7.8	8.0	8.6	0.1	107
Dec.	1974	IV	0.6	8.0	7.0	5.9	1.6	87
Jan.	1975	IV	0.4	12.6	11.6	12.1	1.4	104
Feb.	1975	IV	0.4	9.0	9.2	8.2	0.2	89
Mar.	1975	IV	0.2	15.1	14.5	13.5	0.8	93
Apr.	1975	V	1.1	12.4	12.1	N.AV.	1.4	N.AV.
May	1975	VI	1.6	11.9	11.9	N.AV.	1.6	N.AV.
June	1975	VI	1.2	13.8	13.3	N.AV.	1.7	N.AV.
July	1975	VI	1.8	14.7	14.7	N.AV.	1.8	N.AV.

* Col. III = Col. I + Col. II - Col. IIIa

$$\text{** Col. V} = \frac{\text{Col. IIIa} + \text{Col. IV}}{\text{Col. I} + \text{Col. II}}$$

TABLE 27. MONTHLY TREATMENT SYSTEM PHOSPHORUS BALANCES (EXCLUDING SURGE TANK AND CHLORINE CONTACT CHAMBER) FOR FLOW SHEETS IV, V, AND VI

			P-Balance in terms of mg of P applicable to 1 liter of average monthly flow				P-Balance in terms of percent P introduced into plant by surge tank effluent			
Month	Year	Flow Sheet	Contained in Surge Tank Effluent (mg/ℓ)	Removed in Primary Settler (mg/ℓ)	Removed in Secondary Settler* (mg/ℓ)	Contained in Secondary Effluent (mg/ℓ)	Contained in Surge Tank Effluent (%)	Removed in Primary Settler (%)	Removed in Secondary Settler (%)	Contained in Secondary Effluent (%)
Aug.	1974	IV	11.5	N.AP.	9.5	2.0	100	N.AP.	83	17
Sept.	1974	IV	9.7	N.AP.	8.9	0.8	100	N.AP.	92	8
Oct.	1974	IV	9.9	N.AP.	9.4	0.5	100	N.AP.	95	5
Nov.	1974	IV	9.1	N.AP.	7.5	1.6	100	N.AP.	82	18
Dec	1974	IV	10.3	N.AP.	5.7	4.6	100	N.AP.	55	45
Jan.	1975	IV	9.6	N.AP.	7.6	2.0	100	N.AP.	79	21
Feb.	1975	IV	10.2	N.AP.	7.0	3.2	100	N.AP.	69	31
Mar.	1975	IV	10.6	N.AP.	8.6	2.0	100	N.AP.	81	19
Apr.	1975	V	10.0	1.1	6.2	2.7	100	11	62	27
May	1975	VI	9.2	1.5	5.4	2.3	100	16	59	25
June	1975	VI	9.1	1.9	5.7	1.5	100	21	63	16
July	1975	VI	8.3	2.4	5.0	0.9	100	29	60	11

* By deduction.

TABLE 28. EFFECT OF FERRIC CHLORIDE ADDITION ON PROCESS PHOSPHORUS REMOVAL AND IRON CONCENTRATION OF RETURN SLUDGE: FLOW SHEET IV

Month	Year	Total P in Influent to Aeration Train (mg/ℓ)	Fe^{+++} Dosage (mg/ℓ)	Molar Ratio Fe/P	Fe^{+++} Content of Return Sludge (%)	Total P Removed (%)
Aug.	1974	11.5	12.5	0.60	8.6	83
Sept.	1974	9.7	13.7	0.78	8.4	92
Oct.	1974	9.9	8.7	0.49	5.4	95
Nov.	1974	9.1	7.8	0.47	4.3	82
Dec.	1974	10.3	8.0	0.43	4.1	55
Jan.	1975	9.6	12.6	0.73	5.8	79
Feb.	1975	10.2	9.0	0.48	3.7	69
Mar.	1975	10.6	15.1	0.79	6.7	81

TABLE 29. FLOW-WEIGHTED, MONTHLY-AVERAGE CHLORINE CONTACT CHAMBER EFFLUENT CHARACTERISTICS FOR FLOW SHEET IV*

Parameter		Aug. 1974	Sept. 1974	Oct. 1974	Nov. 1974	Dec. 1974#	Jan. 1975	Feb. 1974	Mar. 1975
Daily Flow, Surge Tank Effluent	(m^3)	534	583	451	511	477#	466	432	458
Daily Flow, Surge Tank Effluent	(mil gal)	(0.141)	(0.154)	(0.119)	(0.135)	(0.126)#	(0.123)	(0.114)	(0.121)
Temperature of Chl. Cont. Chamber Effluent	(C)	N.AV.	22.2	20.7	18.9	15.6#	15.8	15.0	17.2
Temperature of Air	(C)	20.5	14.3	5.8	3.9	-1.1	-2.6	-2.2	-0.6
DO	(mg/ℓ)	N.AV.	3.8	4.5	4.9	5.2#	4.9	4.3	4.9
pH		7.0	7.1	7.0	7.1	7.2#	7.0	7.2	7.1
SS	(mg/ℓ)	4.0	6	10	18	60#	16	16	11
COD	(mg/ℓ)	N.AV.	33	33	44	96#	46	56	53
Diss. COD	(mg/ℓ)	N.AV.	28	28	29	N.AV.	37	36	47
% Diss. COD	(%)	N.AV.	85	85	66	N.AV.	80	64	88
BOD_5**	(mg/ℓ)	N.AV.	3	3	4	26#	9	12	9
Diss. BOD_5**	(mg/ℓ)	N.AV.	3	2	2	N.AV.	5	4	7
% Diss. BOD_5	(%)	N.AV.	100	67	50	N.AV.	56	33	78
NH_4^+-N	(mg/ℓ)	0.3	0.5	0.1	1.6	3.1#	0.7	1.1	2.4
$NO_2^-+NO_3^-$-N	(mg/ℓ)	1.5	1.3	1.7	1.2	2.2#	1.2	1.0	1.6
Diss. Orthophosphorus	(mg/ℓ)	1.9	0.4	0.3	0.7	2.1#	1.0	2.1	1.2

December data are of little significance because plant was operated as a primary plant for part of the month due to breakdown of final settler equipment.

* Average chlorine dose was 6.2 mg/ℓ.

** BOD_5 samples were seeded before analysis.

TABLE 30. FLOW-WEIGHTED, MONTHLY-AVERAGE PRIMARY SETTLER EFFLUENT CHARACTERISTICS FOR FLOW SHEETS V AND VI

Parameter		1975 April	1975 May	1975 June	1975 July
Flow Sheet		V	VI	VI	VI
Daily Flow, Surge Tank Effluent	(m^3)	473	477	572	633
Daily Flow, Surge Tank Effluent	(mil gal)	(0.125)	(0.126)	(0.151)	(0.167)
Temperature of Prim. Sett. Effluent	(C)	19.5	22.4	24.1	26.9
SS	(mg/ℓ)	164	157	184	138
COD	(mg/ℓ)	356	321	298	245
BOD_5	(mg/ℓ)	155	127	110	88
NH_4^+-N	(mg/ℓ)	14.7	16.0	14.7	11.4
TKN	(mg/ℓ)	27.3	27.3	25.3	20.7
NH_4^+-N as % of TKN		54	59	58	55
NO_2^-+NO_3^--N	(mg/ℓ)	0.09	0.07	0.14	0.17
Total Nitrogen	(mg/ℓ)	27.4	27.4	25.4	20.9
TKN as % of COD		8	9	8	8
Ratio COD/TKN		13.0	11.7	11.8	11.8
Diss. Orthophosphorus	(mg/ℓ)	5.0	2.7	2.9	2.2
Diss. Phosphorus	(mg/ℓ)	4.9	2.8	2.6	1.9
Total Phosphorus	(mg/ℓ)	8.9	7.7	7.2	5.9
DO	(mg/ℓ)	0.3	0.3	0.3	0.2

TABLE 31. FLOW-WEIGHTED, MONTHLY-AVERAGE CHLORINE CONTACT CHAMBER EFFLUENT CHARACTERISTICS FOR FLOW SHEETS V* AND VI**

		April 1975	May 1975	June 1975	July 1975
Flow Sheet		V	VI	VI	VI
Daily Flow, Surge Tank Effluent	(m^3)	473	477	572	633
Daily Flow, Surge Tank Effluent	(mil gal)	(0.125)	(0.126)	(0.151)	(0.167)
Temperature of Chl. Cont. Chamber Effluent	(C)	17.2	21.9	23.8	26.8
Temperature of Air	(C)	4.6	17.5	19.9	23.5
DO	(mg/ℓ)	4.2	5.2	5.6	4.5
pH		7.2	7.3	7.2	7.1
SS	(mg/ℓ)	9	9	6	4
COD	(mg/ℓ)	61	55	50	46
Diss. COD	(mg/ℓ)	51	49	44	45
% Diss. COD		84	87	88	98
BOD_5 †	(mg/ℓ)	5	7	2	2
Diss. BOD_5 †	(mg/ℓ)	3	4	1	2
% Diss. BOD_5		60	57	50	100
NH_4^+-N	(mg/ℓ)	1.8	0.6	0.2	0.1
NO_2^-+NO_3^--N	(mg/ℓ)	3.4	3.6	4.2	3.4
Diss. Orthophosphorus	(mg/ℓ)	2.1	1.9	1.2	0.7

* Average chlorine dose for Flow Sheet V was 6.2 mg/ℓ.

** Average chlorine dose for Flow Sheet VI was 5.4 mg/ℓ.

† BOD_5 samples were seeded before analysis.

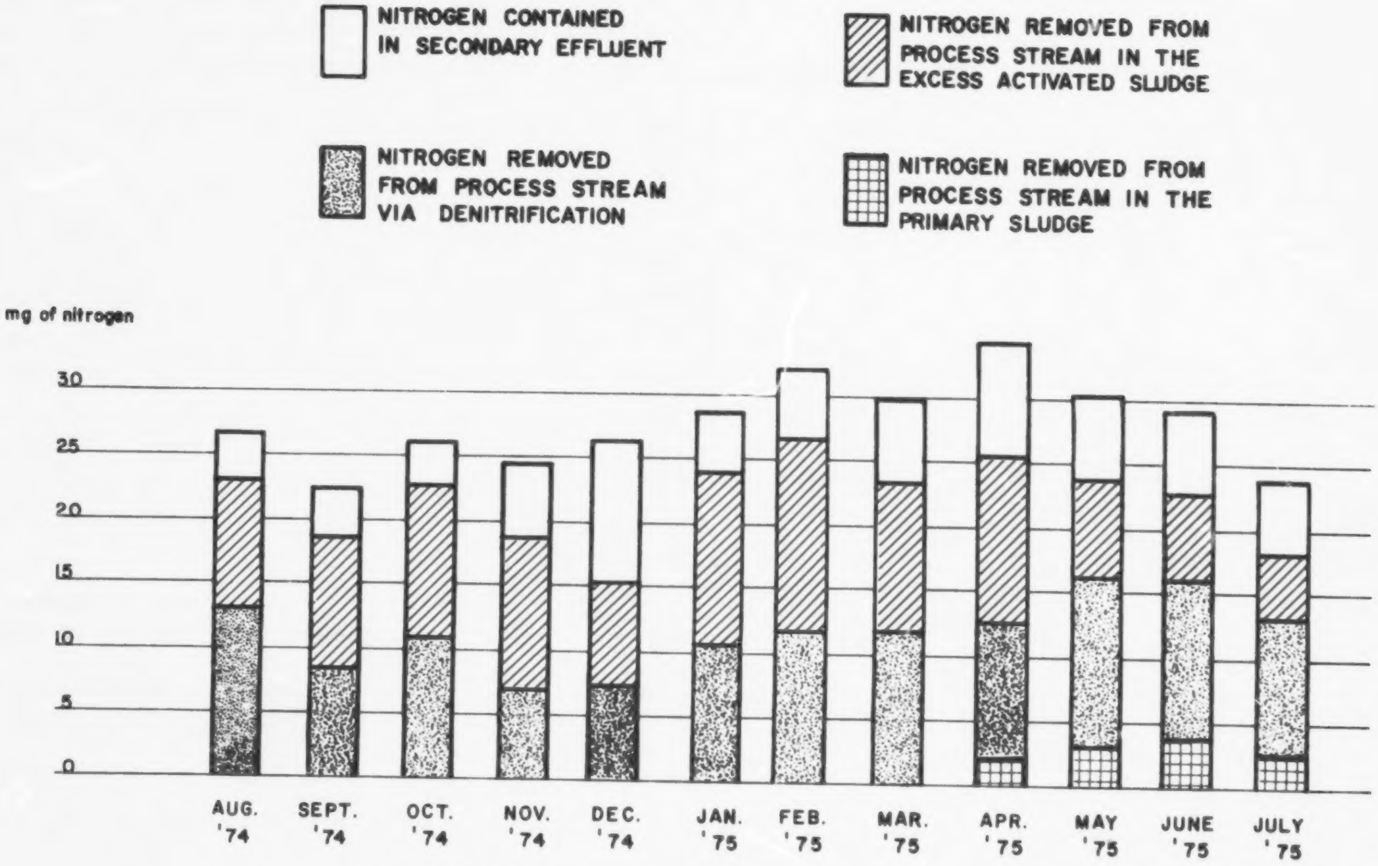


Figure 22. Monthly nitrogen balance bar diagrams for Flow Sheets IV, V, and VI.

The same computational procedure was applied to temperatures and MLSS and RSSS concentrations.

The volatility of MLSS and RSSS in an A.S. system using iron and aluminum salts for phosphorus removal cannot readily be compared with the volatility of sludge occurring in a system not using this procedure. This was commented upon by Mulbarger (22). To permit better comparison of the volatilities achieved, the column "% Vol. Adjusted" was provided in Table 21. It was assumed that all iron in the sludge was present in the form of $Fe(OH)_3$ and that the molecular weight of $Fe(OH)_3$ is equal to 1.91 times the molecular weight of Fe (107/56 = 1.91).

The following equation was used for computing the values in the column headed "% Vol. Adjusted":

$$\text{\% Vol. Adjusted} = \left(\frac{\text{\% Vol.}}{100 - \text{\% Fe} \times 1.91}\right) 100 \qquad (39)$$

where: % Vol. = percent volatile solids determined

% Fe = percent Fe^{3+} found in (total) MLSS or RSSS.

In other words, the volatile solids mass is related to the total solids mass minus the mass of the assumed iron compound therein.

OPERATING RESULTS: SURGE TANK

The effect of the surge tank on raw sewage characteristics was shown previously in Table 7. Sewage detention times in the surge tank are presented in Table 15. Dissolved BOD_5 decreased by 60 mg/ℓ; particulate BOD_5 increased by 5 mg/ℓ. The change in BOD_5 components indicates biological oxidation of dissolved BOD_5 by dispersed biota, with the emphasis on respiration. Alkalinity increased by 25 mg/ℓ.

Over extended periods, inflow to and outflow (outpumping) from an inline surge tank must balance each other. This can be arranged in basically two ways:

a) By adjusting the outpumping rate once or twice a day. At the same time, the sludge recycle rate must be adjusted if that rate is generally kept at a moderate percentage of the inflow. This method of flow control requires constant attention on the part of the plant operator.

b) By adjusting the outpumping rate only to the seasonal and perhaps to the weekend fluctuations of inflow and relying otherwise on the low-level cutout mechanism of the discharge pumps. Controls should be provided that would stop sludge wasting and preferably also effluent chlorination during cutout periods. The sludge blanket in the final settler must generally be kept low under this mode of operation.

By and large, operation of the surge tank was trouble free. A certain amount of disturbances were caused by rags that tended to "ball up" from time-to-time in the submersible torque-flow pump and lodge between the housing and the impeller. Rags also tended to collect around the columns supporting the motor of the floating aerator. Once a block of wood jammed the impeller of the aerator. The surge tank had to be dewatered several times in order to remove 6-cm (2-1/2 in.) diameter can tops from the 10-cm (4 in.) pipeline that connects the center well of the surge tank with the pump pit.

OPERATING RESULTS: FLOW SHEET IV

This flow sheet, shown diagrammatically in Figure 19 and characterized by the omission of primary settling and the utilization of a single anoxic zone in the aeration train comprising Cells 5, 6, 7/8, 9, and 10, was evaluated from August 1974 through March 1975. The aerator in the surge tank was on a 2 min "on", 18 min "off" cycle. Several modes of ferric chloride dosing were employed. This was the first flow sheet under which a rather complete sampling and analytical program was in force.

Operating conditions and performance results for Flow Sheet IV are reflected in Tables 14-29 and Figure 22. Certain aspects of nitrogen removal and sludge production are discussed in two later subsections.

BOD_5, Suspended Solids, and Nitrogen (N) Removal

Provided that a system has sufficient N removal capacity via sludge synthesis and denitrification, N removal efficiency will depend on the flocculating properties of the sludge and will parallel SS and BOD_5 removals. Poor flocculating properties will be reflected in high SS in the effluent, paralleled by high BOD_5 and TKN concentrations. This was observed on this project and may be seen by examining the respective tables.

It appears that a change in the nature of the wastewater being treated occurred during December 1974, resulting in increased final effluent BOD_5, SS, and TKN. The effluent became "milky" again, thus creating a very unsightly appearance in the final settler. The fine SS could be filtered out by a 0.45 μm membrane filter, but not by a glass fiber filter. This led to numerous attempts to remedy the situation, among others the relocation of the $FeCl_3$ dosing point from the surge tank effluent to the final settler influent. That measure improved the situation somewhat, but not to a satisfactory level. Consistently low SS concentrations in the final effluent were established only under Flow Sheet VI by adding 8 mg/ℓ of Fe^{3+} in the form of $FeCl_3$ to the influent of the primary settler.

It was theorized that some new constituent in the process water had an unfavorable effect on flocculation. Nitrification and denitrification were not seriously affected, but phosphorus removal suffered considerably. The deterioration of SS removals was not caused by the somewhat lower winter temperatures of the process water; such deterioration was not observed in the preceding winter season.

The occurrence of the "milky" effluent was accompanied by heavy foaming in the aeration train and also in the mixing chamber of the CCC. In general, a poor effluent, in terms of BOD and SS, was associated with the heavy foaming.

Ferric Chloride Addition and Phosphorus Removal

Ferric chloride dosage, in terms of mg/ℓ Fe^{3+} and Fe/P molar ratio, is shown in Table 28. The Fe/P molar ratio varied between 0.43 and 0.79. Up to September 20, ferric chloride was dosed to the effluent of the surge tank. On that day, the dosing point was relocated to the effluent of the aeration train. This brought about a drastic decrease in $FeCl_3$ requirements and explains the low $FeCl_3$ dosages in October, November, and December. On February 7, $FeCl_3$ addition was discontinued for 2 wk because Fe appeared in the effluent and because the ML assumed an unusual color. Ferric chloride addition was reinstituted on February 28, this time again to the surge tank effluent.

Disregarding December*, satisfactory phosphorus removal (i.e., 1 mg/ℓ final effluent P, or less) was achieved only in September and October. In addition, 80 percent or better P removal was also achieved in August, September, and March. During February, ferric chloride feeding had been discontinued for 20 days. This explains the poor P removal results in February.

In general, P removal efficiency paralleled SS removal efficiency. The same unknown wastewater constituent that adversely affected SS removal also affected P removal unfavorably.

Efficient P removal by metal addition to the A.S. process is predicated on the general soundness of the process, as observable in SS removal. When clarification efficiency is impaired due to wastewater characteristics, increased metal dosage generally will not help.

OPERATING RESULTS: FLOW SHEET V

Except for changes in the aeration train aeration pattern, the mode of ferric chloride addition, and the pattern of surge tank aeration, Flow Sheet V, as seen schematically in Figure 20, was identical to Flow Sheet II. The aeration pattern utilized was the same as employed with Flow Sheet IV. Ferric chloride was dosed in a "split" fashion, 4 mg/ℓ Fe^{3+} to the ML leaving the aeration train, the rest to the primary effluent. The surge tank aerator was operated in the cyclic pattern adopted in Flow Sheet IV, rather than the full "on"-time mode used with Flow Sheet II.

Since Flow Sheet II, several mechanical modifications had been made to the primary settler: the sludge wasting appurtenances had been improved and a 25-mm x 25-mm (1-in. x 1-in.) coarse mesh screen had been installed in the effluent trough.

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* The December results are not significant because the secondary plant was in operation only part of the month.

Operating conditions and performance results for this flow sheet are presented in Tables 14-27, 30, and 31 and Figure 22. Certain aspects of nitrogen removal and sludge production are discussed in a later subsection.

Final effluent BOD_5, dissolved BOD_5, and SS concentrations were similar to the average values achieved during Phase IV. However, average effluent COD and dissolved COD were 35 and 49 percent higher, respectively. Also, total N removal efficiency deteriorated to unsatisfactory levels and the Phase V run was discontinued at the end of April 1975 after 1 mo of operation. It is possible that the operating period was too short to allow performance to be optimized.

OPERATING RESULTS: FLOW SHEET VI

As depicted schematically in Figure 21, Flow Sheet VI was characterized by the use of the primary settler and split ferric chloride addition (8 mg/ℓ Fe^{3+} to the influent of the primary settler and 5-6 mg/ℓ Fe^{3+} to the aeration tank effluent). Experimentation with the aeration train aeration pattern, intended to improve denitrification, was carried out during this final phase which lasted 3 mo (May-July 1975). The cyclic 2 min "on", 18 min "off" operating mode for the surge tank aerator was continued.

Operating conditions and performance results for this flow sheet are reflected by the data in Tables 14-27, 30, and 31 and Figure 22. Certain aspects of N removal and sludge production are discussed in two later subsections.

The increase in Fe^{3+} dosing rate and the forward (upstream) movement in one of the split Fe^{3+} dosing points were based on the idea that the upstream addition of ferric chloride, an acidic compound, would help to alleviate the detrimental effects of the unknown institutional contaminant that inhibited complete flocculation in the A.S. process, thus resulting in periodic solids removal problems. It was assumed that these contaminants originated either in the kitchen or the laundry. The upstream dosing would allow the ferric chloride to react with the process water outside of the A.S. tank. The change in the ferric chloride dosing procedure was not related to phosphorus removal considerations. Experimentation with the aeration pattern was initiated only to counteract anticipated effects of the change in the ferric chloride dosing scheme on N removal.

Operating Conditions

The average hydraulic load on the aeration system was somewhat higher, 575 m^3/day (0.15 mgd) vs. 506 m^3/day (0.13 mgd), in Phase VI than in Phase IV. The average process water temperature was 25 C.

Table 32 compares contaminant concentrations in the influent to the aeration tank for Flow Sheets IV and VI. Aeration system influent BOD_5 and SS concentrations for Flow Sheet VI were only 55 and 65 percent, respectively, of those for Flow Sheet IV. Influent TKN concentration, however, changed little.

The decrease in influent BOD_5 and SS concentrations had the effect of halving sludge production in terms of the mass of solids produced in the A.S. system

TABLE 32. AVERAGE CONTAMINENT LOAD ON AERATION TRAIN DURING FLOW SHEET IV AND VI OPERATIONS

Parameter	Flow Sheet IV* (mg/ℓ)	Flow Sheet VI** (mg/ℓ)
SS	245	160
COD	422	287
BOD_5	196	107
NH_4^+-N	12.2	13.8
TKN	27.0	24.1
Total Phosphorus	10.1	6.9
Diss. Orthophosphorus	4.9	2.6

* Surge Tank Effluent.
** Primary Settler Effluent.

per liter of sewage treated as seen in Table 22. Solids produced per mg of BOD_5 removed, on the other hand, did not change significantly. The volatility of the sludge did not increase in Phase VI as one would have expected considering the removal of inert solids in the primary settler. The lack of increase in ML volatility may have been due to the higher Fe^{3+} content of the sludge. The BOD_5 reduction brought about by primary settling plus $FeCl_3$ addition in Phase VI was 28, 30, and 38 percent, respectively, in the months of May, June, and July of 1975.

The effect of primary settling assisted by metal addition on sludge production and SRT is shown in the comparison of Flow Sheet IV (no primary settling) and Flow Sheet VI (primary settling with $FeCl_3$ addition) data in Table 33 below.

Aeration Pattern*

Several measures were taken with the goal of improving aeration train N removal. These included the following:

1) Cell 11 was converted to anoxic operation to extend the anoxic detention time of the aeration train. This resulted in the retention of only one cell, Cell 12, for aerobic polishing following denitrification. This attempt at extending the anoxic portion of the aeration train had to be abandoned. One cell at the end of the train was not sufficient to nitrify all the ammonia nitrogen released during the anoxic part of the cycle.

2) Cell 4 was also converted to an anoxic operating mode, again to increase the denitrification period. It was found that the first three cells of the train nitrified all the ammonia nitrogen in the primary effluent. This was a beneficial, permanent change.

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* See Table 14 for plant air flows and aeration cell DO's.

TABLE 33. COMPARISON OF SLUDGE PRODUCTION AND ORGANIC LOAD DATA FOR FLOW SHEETS IV AND VI

Parameter	Flow Sheet IV	Flow Sheet VI
Solids Production (mg of EAS and final effluent SS produced per mg of BOD_5 removed)	1.15	1.20
Solids Production (mg of EAS and final effluent SS produced per liter of sewage treated)	211	123
Liquid EAS Production (liter per liter of sewage treated)	0.017	0.008
Solids Residence Time Based on MLSS (days)	4.7	8.0
Daily BOD_5 Load on MLSS (kg/kg)	0.20	0.11

3) An alternating pattern of aeration-mixing was tried in July 1974 with the idea of promoting some substrate nitrate respiration to occur. After initially promising results, the success of this trial was not decisive. The attempt was abandoned after 1 mo. DO in the aeration cells must be rigidly controlled to assure good denitrification in the anoxic cells. Too high a DO concentration in an aerobic cell followed by an anoxic cell will yield an unfavorably high DO in the anoxic cell and adversely affect denitrification in that cell. Also, it appeared that the aerobic-anoxic pattern had a deteriorating effect on denitrifying ability, decreasing the fraction of facultative anaerobes present. The alternating pattern used was as follows:

Cell 1	Cell 2	Cell 3	Cell 4	Cell 5	Cell 6	Cell 7/8
aerobic	anoxic	aerobic	anoxic	aerobic	anoxic	anoxic

Cell 7/8	Cell 9	Cell 10	Cell 11	Cell 12
anoxic	aerobic	anoxic	anoxic	aerobic

4) An attempt was made to increase sludge production in the aeration train (and associated N removal) by keeping DO concentrations in the aerobic cells low. This experiment also had no readily discernible success.

Treatment Results

Final effluent BOD_5 and SS residuals decreased to their lowest sustained levels of the project, 7 and 8 mg/ℓ, respectively, on the average. Effluent

COD increased slightly, however, compared to Flow Sheet IV, 55 mg/ℓ vs. 49 mg/ℓ, due to a 40 percent increase in residual dissolved COD. Total N removal also deteriorated slightly to approximately 2 percent less than the 80 percent target. The denitrification rate stayed about the same as under Flow Sheet IV, but N removal via incorporation in EAS fell sharply due to the lower F/M loading and the resulting lower total EAS production. This decrease in N removal via EAS production was not fully compensated for by the additional N removed in the primary sludge.

A second, very pronounced, change in process behavior was the increase in the rate of nitrification. Nitrification remained complete, despite the facts that, compared to Flow Sheet IV, 13.6 percent more unoxidized N entered the aeration train on a mass loading basis and the initial (as opposed to polishing) nitrification step was accomplished in three cells instead of four.

Another new phenomenon observed with this flow sheet was the sizeable release of ammonia nitrogen during the anoxic period. Obviously what occurred here was a shift from adsorbed carbon nitrate respiration to true ENR.

Conclusions

The reaction time provided by Flow Sheet VI was not sufficient, under the prevailing operating conditions, to provide adequate denitrification of the increased nitrate nitrogen load on the anoxic zone of the aeration system. An additional ML detention time of at least 1 hr would have been required for ENR to be carried to completion. The increased CRT employed with this flow sheet (8-9 days) significantly enhanced SS and BOD_5 removals.

NITROGEN PROFILES: FLOW SHEETS I, IV, AND VI

From data collected in aeration train profile sampling (see Appendix Tables B-1 to B-7), the N profiles of Figures 23-29 were drawn. These graphs show, in an idealized way, the concentration changes of the various forms of dissolved N for a slug of process water traversing the aeration train. Actually, these changes appear to occur in the form of concentration jumps due to the CM nature of the individual compartments. For each cell depicted, two values of N concentrations are shown, the influent values at the left side, the effluent values at the right side. The influent values were obtained by analyzing the sample of the preceding cell, the effluent values by analyzing the cell sample itself. For the profiles of June 12, 1975 (Figure 29 - Flow Sheet VI), the first sample was taken from a mixing chamber where primary effluent was admixed to return sludge. These profiles, therefore, reflect the reactions in the entire aeration train. For the other profile graphs, the first sample was taken from the first cells; reactions occurring in the first cell are, therefore, not depicted in these other profile graphs. Soluble, organically bound N is represented by the distance between the TKN and NH_4^+-N curves.

The bar diagrams of Figures 30-33 indicate via the use of graphic symbols changes in NH_4^+-N and $NO_2^-+NO_3^-$-N concentrations for the individual cells, depicted as nitrification, denitrification, uptake by the biomass, release by the biomass, and ammonification.

MLSS: 3300 mg/ℓ
MLVSS: 2200 mg/ℓ

Flow-Through-Time:
29 min per cell
(except Cell 7/8)

ML Temperature:
20 C est.

Primary Settler
Effluent Treated

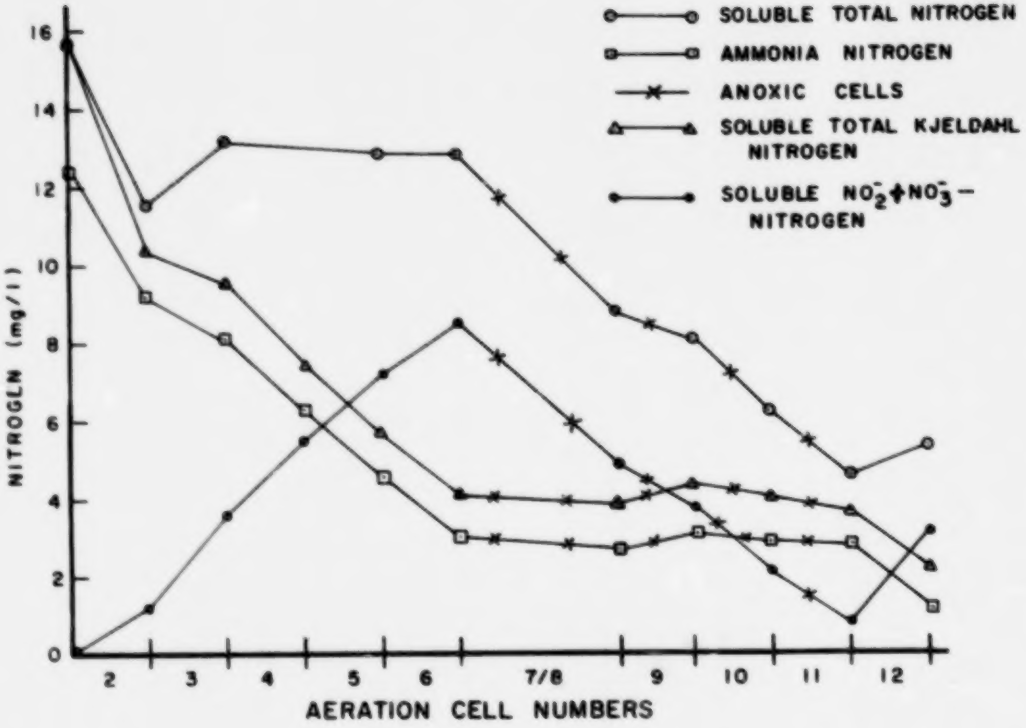

Figure 23. Flow Sheet I nitrogen profiles (11/14/73).

MLSS: 3600 mg/ℓ
MLVSS: 2200 mg/ℓ

Flow-Through-Time:
29 min per cell
(except Cell 7/8)

ML Temperature:
20 C est.

Primary Settler
Effluent Treated

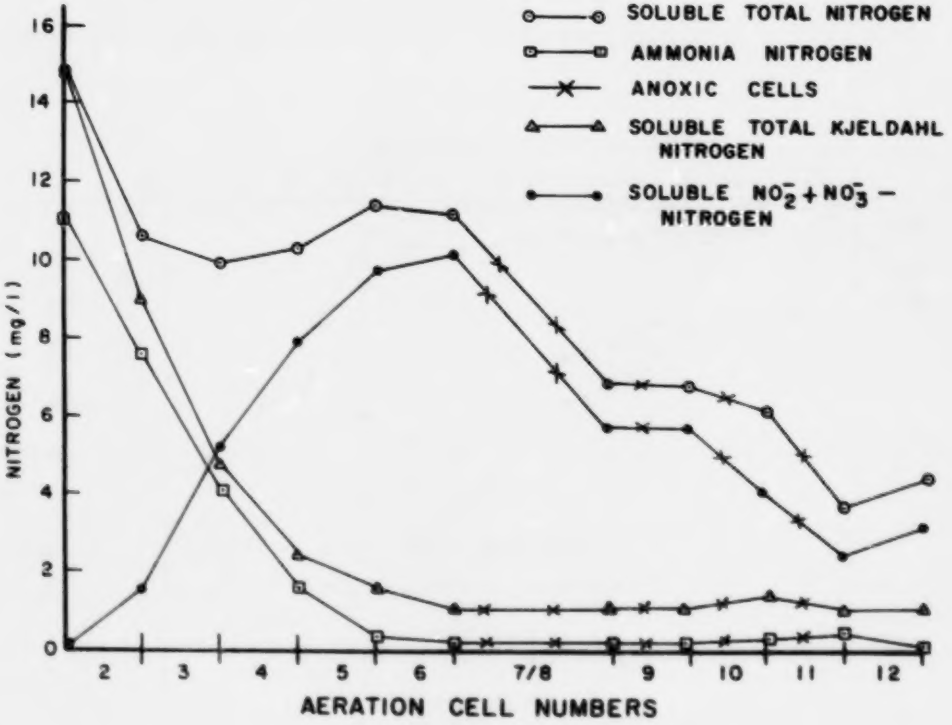

Figure 24. Flow Sheet I nitrogen profiles (11/27/73).

MLSS: 4000 mg/ℓ
MLVSS: 2500 mg/ℓ

Flow-Through-Time:
29 min per cell
(except Cell 7/8)

ML Temperature:
18 C est.

Primary Settler
Effluent Treated:
COD 360 mg/ℓ
TKN 24.6 mg/ℓ
Diss. TKN 12.5 mg/ℓ

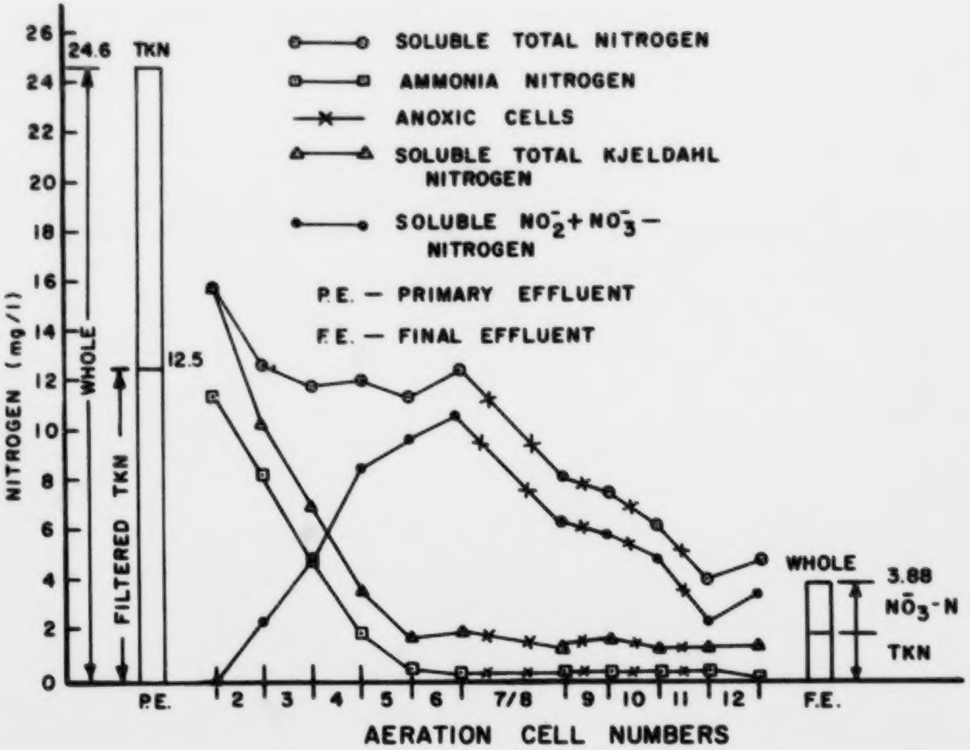

Figure 25. Flow Sheet I nitrogen profiles (12/5/73).

MLSS: 2800 mg/ℓ
MLVSS: 1800 mg/ℓ

ML Flow-Through-Time:
26 min per cell
(except Cell 7/8)

ML Temperature:
23 C

Surge Tank
Effluent Treated:
TKN 26.6 mg/ℓ

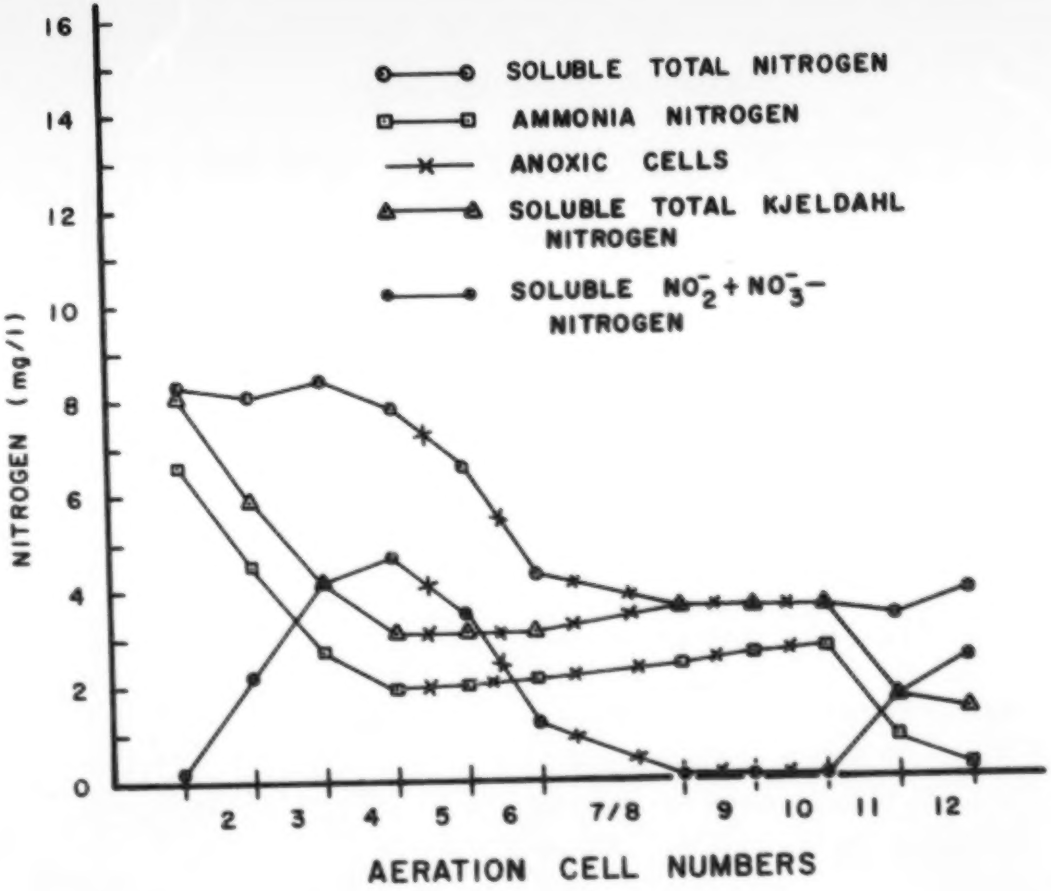

Figure 26. Flow Sheet IV nitrogen profiles (9/18/74).

MLSS: 3900 mg/ℓ
MLVSS: 2600 mg/ℓ

Flow-Through-Time:
26.5 min per cell
(except Cell 7/8)

ML Temperature:
22 C

Surge Tank
Effluent Treated:
TKN 30.2 mg/ℓ
Diss. TKN 17.1 mg/ℓ
NH_4^+-N 12.0 mg/ℓ

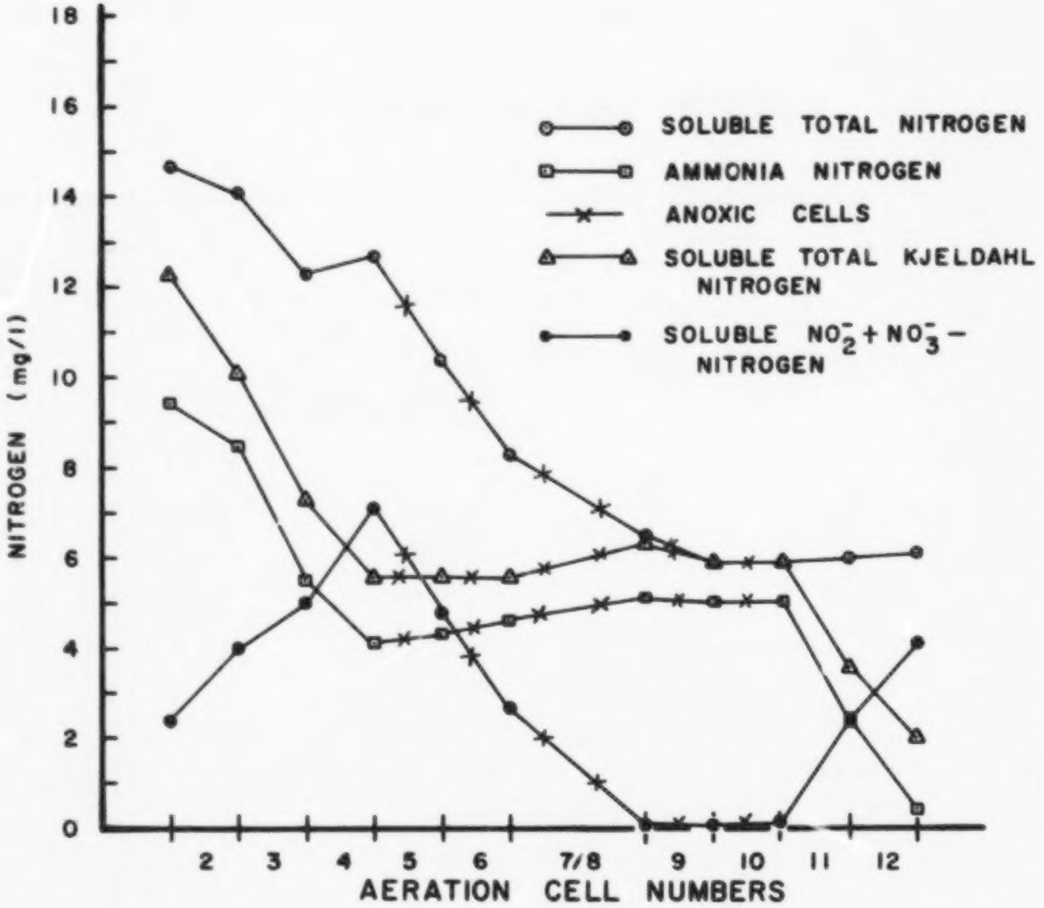

Figure 27. Flow Sheet IV nitrogen profiles (10/10/74).

MLSS: 2700 mg/ℓ
MLVSS: 1600 mg/ℓ

Flow-Through-Time:
25 min per cell
(except Cell 7/8)

ML Temperature:
20 C

Surge Tank
Effluent Treated:
TKN 25.5 mg/ℓ
Diss. TKN 17.9 mg/ℓ
NH_4^+-N 11.4 mg/ℓ

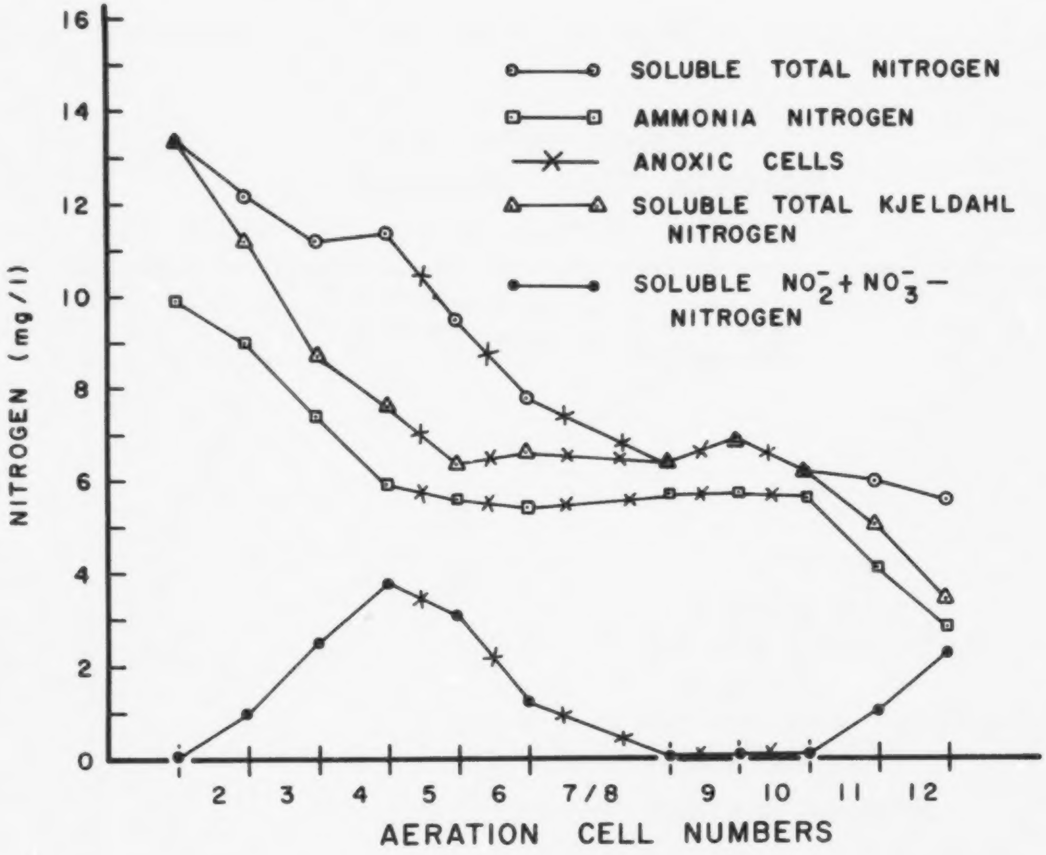

Figure 28. Flow Sheet IV nitrogen profiles (11/14/74).

MLSS: 4150 mg/ℓ
MLVSS: 2600 mg/ℓ

Flow-Through-Time:
28 min per cell
(except Cell 7/8)

ML Temperature:
25 C

Primary Settler
Effluent Treated:
TKN 36.7 mg/ℓ
Diss. TKN 29.7 mg/ℓ
NH_4^+-N 26.0 mg/ℓ

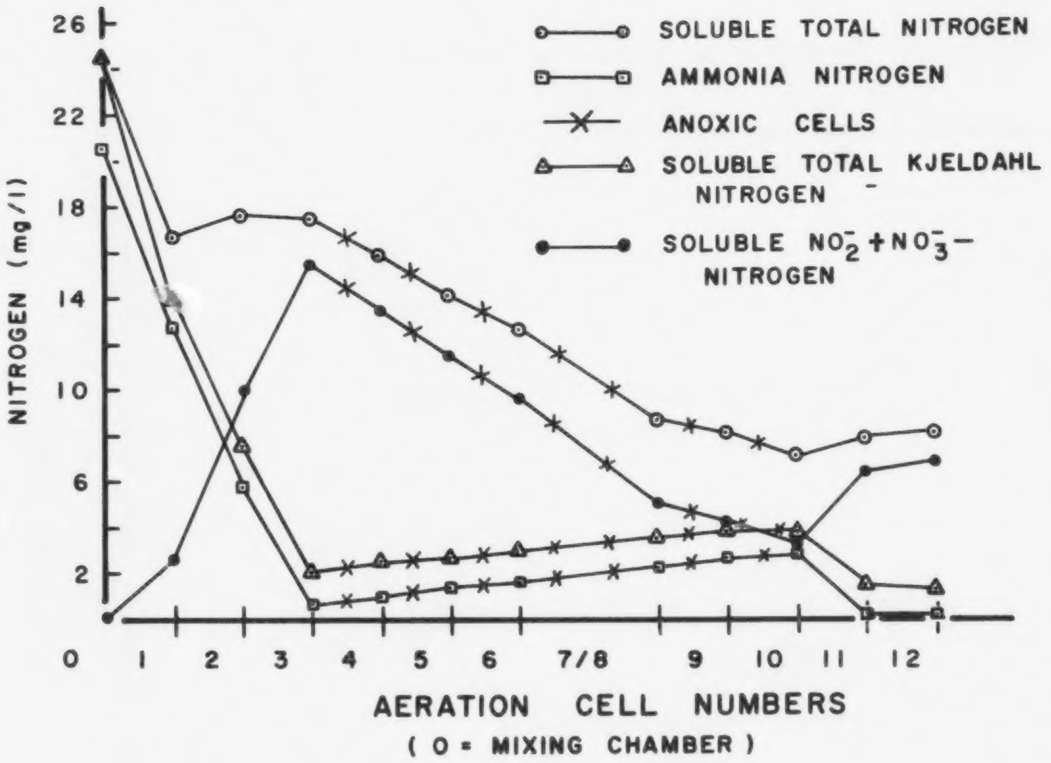

Figure 29. Flow Sheet VI nitrogen profiles (6/12/75).

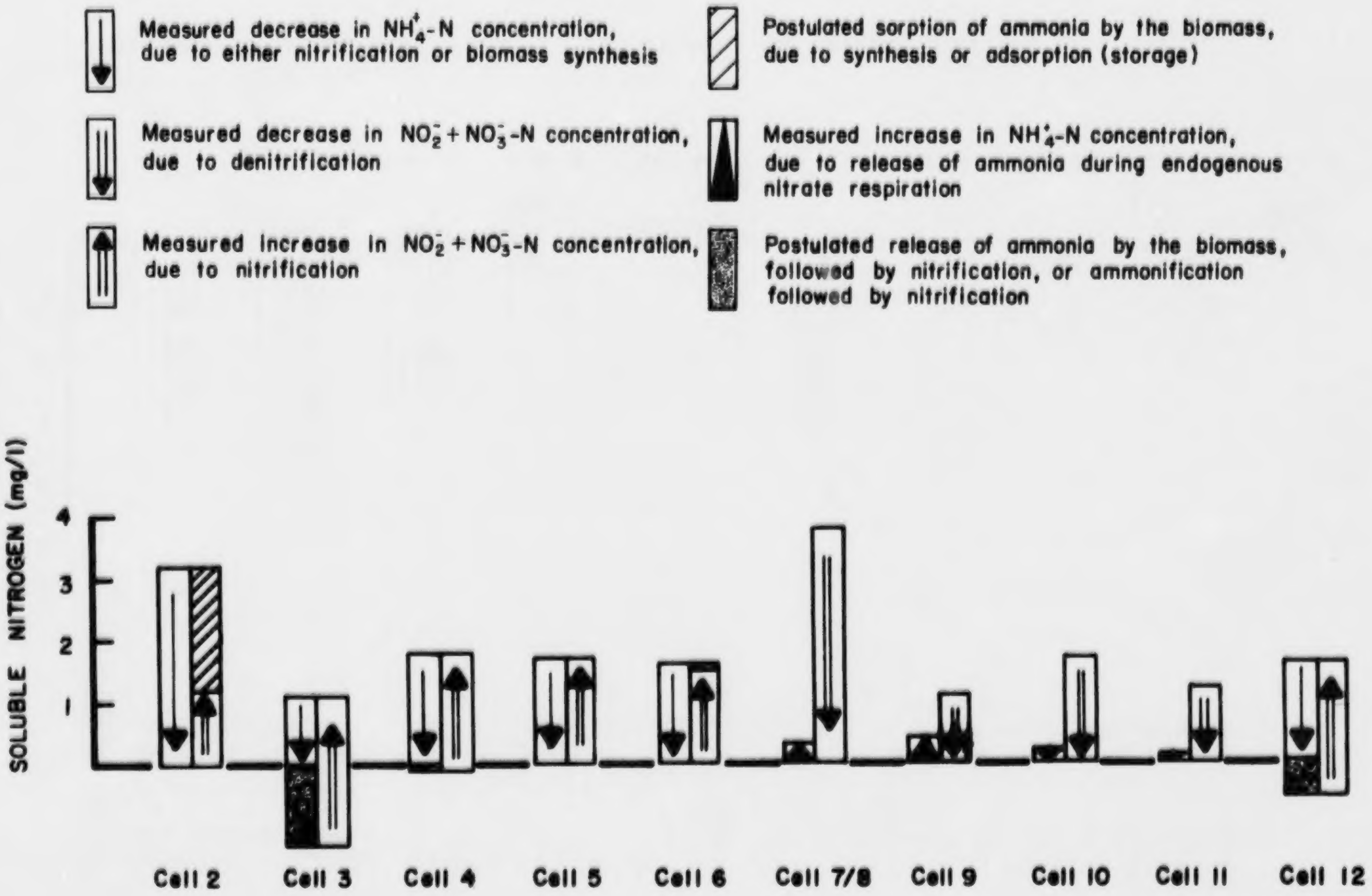

Figure 30. Flow Sheet I ammonia and nitrate bar diagram profiles (11/14/73).

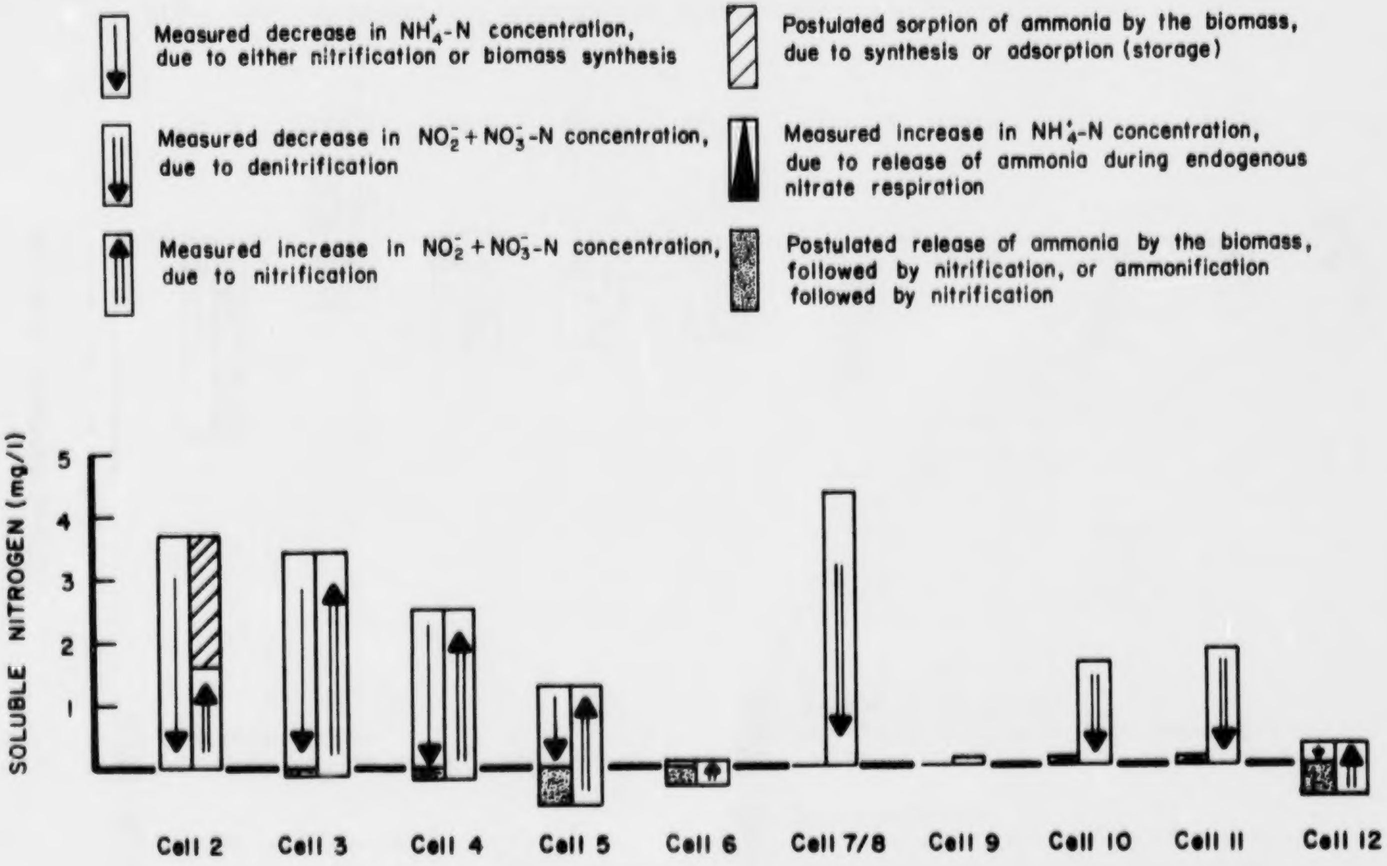

Figure 31. Flow Sheet I ammonia and nitrate bar diagram profiles (11/27/73).

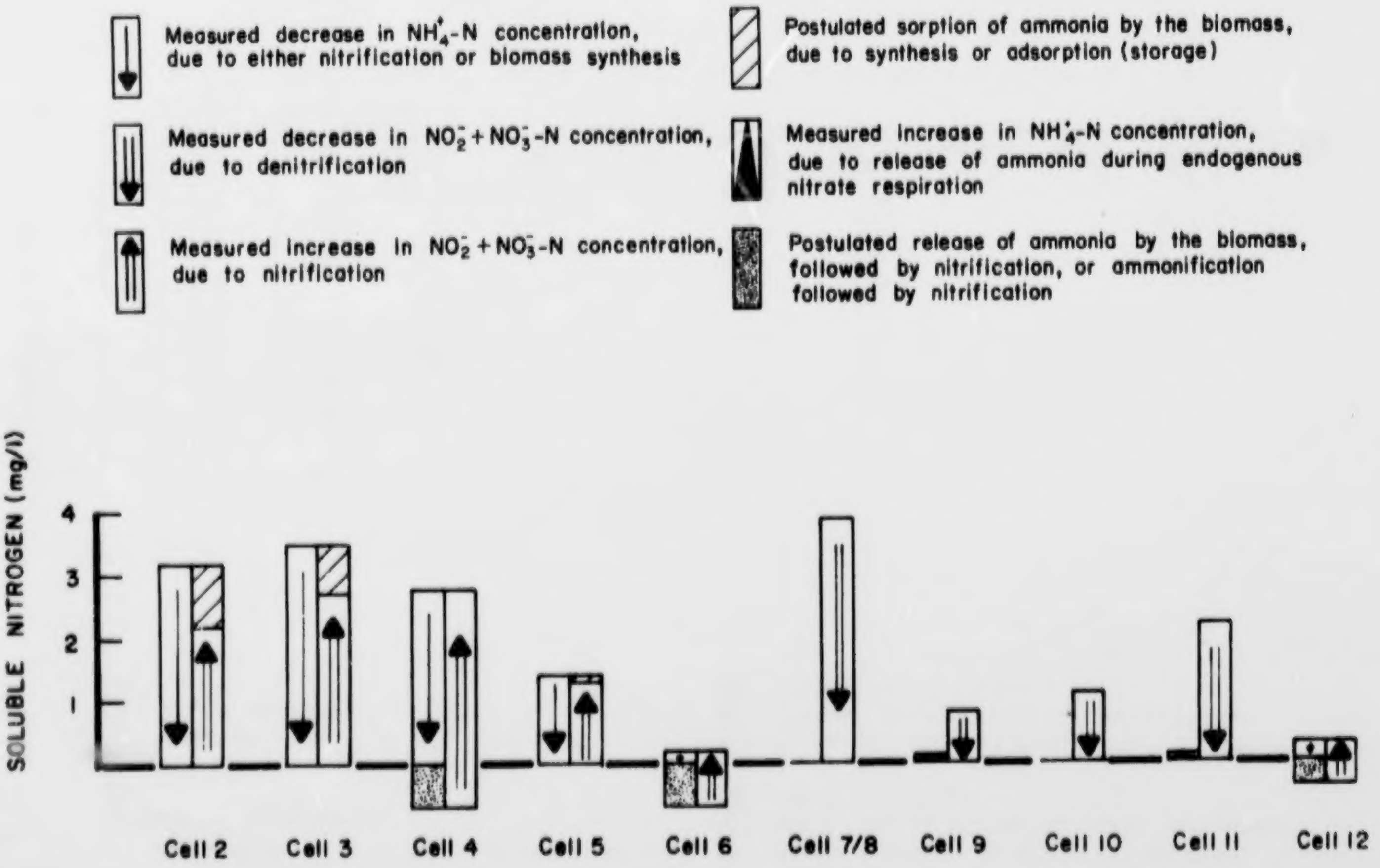

Figure 32. Flow Sheet I ammonia and nitrate bar diagram profiles (12/5/73).

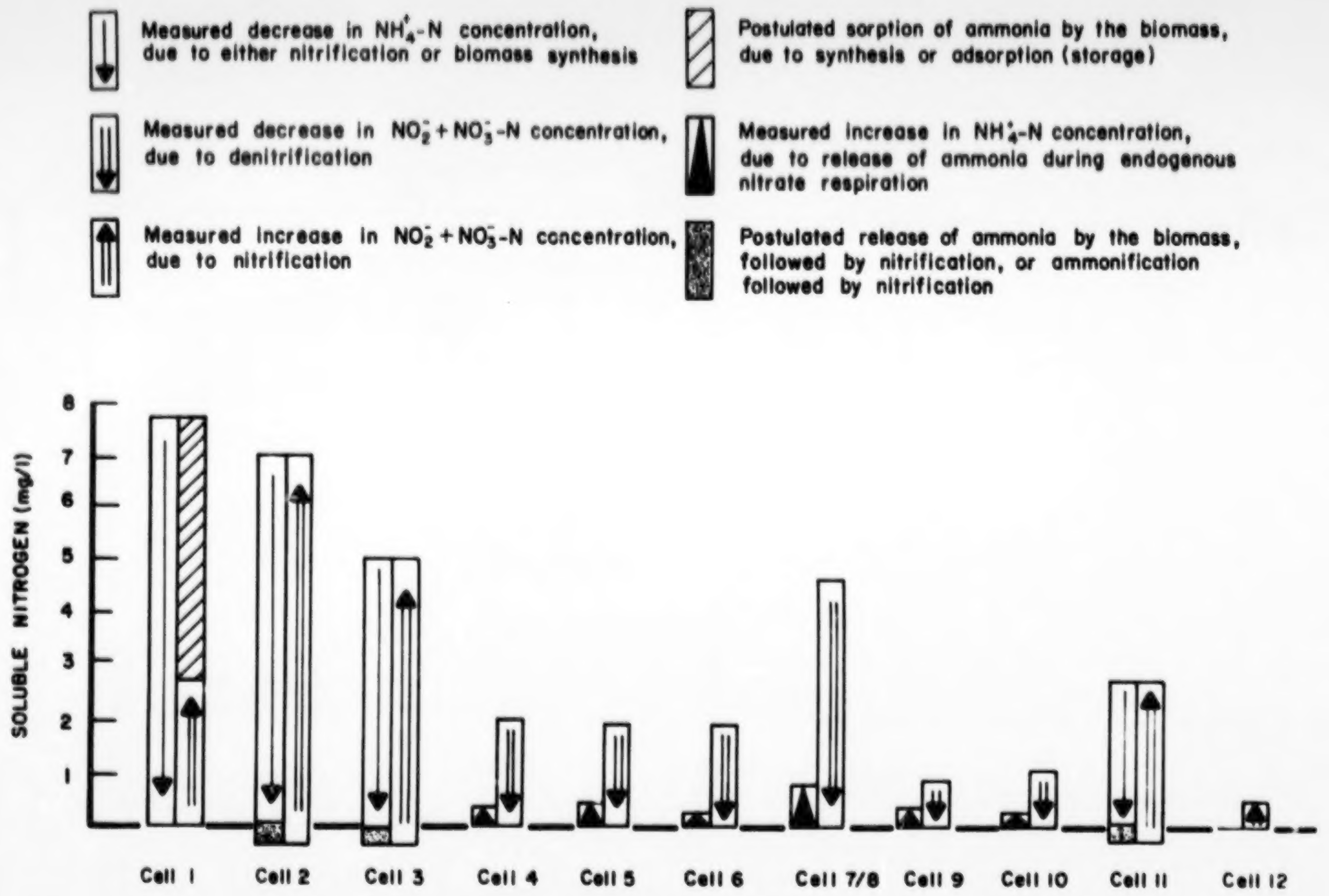

Figure 33. Flow Sheet VI ammonia and nitrate bar diagram profiles (6/12/75).

Shape of Dissolved Total Nitrogen Curves

Figures 23-29 all show at least one period of concentration increase in the first half of the process, as follows:

Figure	23	24	25	26	27	28	29
Flow Sheet	I	I	I	IV	IV	IV	VI
Cell in which increase in dissolved total N occurred	3	4, 5	4, 6	3	4	4	2

This occurrence indicates solubilization of particulate substrate N, mainly to NH_4^+-N. A small amount of organically bound N was also released. The position of this release seems to indicate the efficiency of the aerobic process, an earlier occurrence tending toward greater efficiency. Note that Cell 3 was the second aerobic cell in Flow Sheet I but the third aerobic cell in Flow Sheet IV. This rise in dissolved N was apparently due to the interaction of biota with particulate nitrogenous matter adsorbed by the biota on contact with primary effluent or surge tank effluent.

In all profiles except Figure 28, a second rise of dissolved total N occurs toward the end of the process, due to the release of biomass N as NH_4^+-N. The ammonia nitrogen is immediately nitrified in the following aerobic cells.

Shape of Bar Diagrams

In aerobic cells, the following reaction combinations are apparent:

1) Decrease of NH_4^+-N not matched by nitrification. The difference is taken up by the biomass and is indicated by hatching in the graphs.

2) Nitrification not matched by decrease in NH_4^+-N. The difference is due to NH_4^+-N released by the biomass or to ammonification, both reactions being followed by nitrification. This is indicated by stippled areas in the figures.

In anoxic cells, release of ammonia nitrogen occurs; this is indicated by a black arrow.

Absorption of Ammonia Nitrogen by Biomass vs. Nitrification

The bar diagrams indicate that absorption may be greater than nitrification and occurs in the first one or two aerobic cells. Intensity and forward location of this phenomenon may be indicative of the rate of aerobic synthesis.

Preference of Biota for Organically Bound Nitrogen

The distance between the TKN and NH_4^+-N curves narrows at the beginning of the process. No significant increase in distance occurs further down the cells. Particulate N is utilized very early in the process; most of the particulate matter in the influent is quickly adsorbed by the biomass.

Ammonia Release During Endogenous Nitrate Respiration

Table 34 indicates that the amount of NH_4^+-N released during ENR was generally much smaller than predicted by stoichiometric theory, which calls for a release of 0.25 mg NH_4^+-N/mg N gasified, or 25 percent. The highest release rate encountered was 17 percent. Two profiles for Flow Sheet I exhibit practically no NH_4^+-N release.

To explain the lack of ammonia nitrogen released, one might hypothesize as follows:

a) Adsorbed carbon compounds, or storage carbon not yet assimilated into the biomass, may have been used as the hydrogen donor. This reaction would resemble substrate nitrate respiration and would be accompanied by some sludge production. The release of NH_4^+-N would be minimal.

b) Lack of ammonia nitrogen release may not be real. Nitrification may go on simultaneously with denitrification preventing an increase in NH_4^+-N concentration.

c) The bacterial biomass may contain less nitrogen than indicated by the formula $C_5H_7O_2N$, which has been stipulated by Porges et al. (2).

d) The facultative organisms engaged in nitrate respiration may have utilized some substrate carbon, perhaps deflocculated matter.

The variability of the ammonia nitrogen release observed indicates that more than one of the factors listed above was involved.

Alkalinity Release During Endogenous Nitrate Respiration

The theoretical release of alkalinity during ENR is 4.46 mg/mg NO_3^--N gasified, 3.57 mg of which is due to the reduction of nitrate nitrogen and 0.89 mg to the release of ammonia nitrogen by the biomass. All of the five profiles for which alkalinity data are available (see Table 34) show a lesser release of alkalinity. Three profiles indicate a release near the 3.57 mg level, hinting at the effect of adsorbed carbon respiration; the other two profiles depict a release even below the 3.57 mg level, indicating perhaps nitrification occurring in the anoxic cells.

Alkalinity Destroyed During Nitrification

Table 34 indicates that destruction of alkalinity during nitrification was in fair agreement with the stoichiometric prediction of 7.14 mg/mg NH_4^+-N oxidized.

Rates of Nitrification

The changing rates of nitrification can best be observed by examining the aerobic portions of the NO_2^-+NO_3^--N curves. The NH_4^+-N curves are not suitable for this purpose because the decrease in NH_4^+-N may have been due to sorption by the heterotrophic biomass.

TABLE 34. STOICHIOMETRIC DATA ABSTRACTED FROM NITROGEN PROFILES FOR FLOW SHEETS I, IV, AND VI

Parameter	Date of Profile						
	11/14/73	11/27/73	12/5/73	9/18/74	10/10/74	11/14/74	6/12/75
Flow Sheet	I	I	I	IV	IV	IV	VI
Temperature of Process Water (C)	20	20	18	23	22	20	25
Nitrification (mg/ℓ of NH_4^+-N oxidized)	10.8	10.8	11.2	6.9	8.7	5.8	18.9
Alkalinity Destruction During Nitrification (mg/ℓ)	N.AV.	60.0	82.0	N.AV.	52	45	130
Specific Alkalinity Destruction (mg/mg)	N.AV.	5.6	7.3	N.AV.	6.0	7.8	6.9
Denitrification via EMR (mg/ℓ N gasified)	7.8	7.8	8.0	4.6	7.0	3.7	12.2
Alkalinity Release During Denitrification (mg/ℓ)	N.AV.	20	28	N.AV.	26	8	48
Specific Alkalinity Release (mg/mg)	N.AV.	2.6	3.5	N.AV.	3.7	2.2	3.9
Ammonia Release (Uptake) During EMR (mg/ℓ)	(0.2)	0.2	0.2	0.8	0.9	(0.3)	2.1
Specific Ammonia Release During EMR	Q.R.	0.026	0.025	0.174	0.129	Q.R.	0.173

The rate of nitrification appears to have been inhibited in the beginning phase of the process. The extent of this inhibition or retardation seems to have depended on the overall nitrification efficiency. At high rates of overall efficiency, the retardation was of short duration, not more than 30 min. This retardation seems to be accompanied by intense sorption of NH_4^+-N on part of the heterotrophs. On the other hand, an oxygen deficiency was never noted in any of the aerobic cells. One might therefore theorize that the retardation was due to transport difficulties with respect to ammonia nitrogen, the heterotrophic organisms competing with the n. for NH_4^+-N and gaining the upper hand during a short period of the process. In three of the four bar graphs (Figures 30-33), sorption of NH_4^+-N was greater than nitrification across the first aerobic cell.

Once this period of retardation was overcome, the nitrification curves had the shape of the Monod profiles illustrated previously in Figure 6. The highest rate tended to occur right after the inhibition period; the rate stayed nearly constant down to approximately 2 mg/ℓ NH_4^+-N.

The drastic increase in nitrification rates for the profile of June 12, 1975 (Figure 29) reflects the increase of CRT provided for Flow Sheet VI. This increase is in basic agreement with kinetic theory.

The N profile curves (Figures 23-29) reflect the mg/ℓ/time rates, not these rates referred to 1 mg/ℓ MLVSS. The nitrification rates appear as the slope of the NO_2^-+ NO_3^--N profiles for the aerobic cells. It is logical to inspect the nitrification rates as they appear in the aeration train profiles for it can be assumed that the active biomass does not change appreciably during one reaction period. Also, when comparing different profiles with each other, it should be remembered that the kinetic theory of nitrification in single-sludge A.S. systems indicates that the mass of n. is independent of MLSS concentration; it is dependent only on the ammonia nitrogen load, environmental conditions, and the growth rate imposed on the MLSS. The average and highest nitrification rates associated with the seven aforementioned profiles of Figures 23-29 are summarized in Table 35, on the bases of both mass/volume/time and mass/mass/time.

Rates of Endogenous Nitrate Respiration (Denitrification)

In the aeration train profiles (Figures 23-29), ENR reaction rates are given in terms of mg NO_2^-+NO_3^--N gasified/liter reaction volume vs. time and appear as the slope of the NO_2^-+NO_3^--N profiles for the anoxic cells. In Table 35, rates of denitrification are shown per unit volume of reactor/time and per unit mass VSS/time.

The significance of the ENR rates observed is considerably diminished by the variability of the operating conditions that prevailed. Obviously, ENR rates are affected by the DO concentration of the process water and, if it occurs, by the exhaustion of NO_2^-+NO_3^--N. The latter condition may largely be avoided by proper utilization of compartments. Most cells were kept anoxic by throttling the air flow to the aeration grids located near the bottom of the cells. This type of operation, of course, is not the best way to maintain minimum DO concentrations. With a view to these factors, the highest

TABLE 35. KINETIC DATA ABSTRACTED FROM NITROGEN PROFILES FOR FLOW SHEETS I, IV, AND VI

		Date of Profile						
		11/14/73	11/27/73	12/5/73	9/18/74	10/10/74	11/14/74	6/12/75
Flow Sheet		I	I	I	IV+	IV+	IV+	VI
Temperature of Process Water* (C)		20	20	18	23	22	20	25
Nitrification Rate† (mg NH_4^+-N oxidized/ℓ reactor volume/hr)	Highest	5.0	7.2	7.2	4.6	4.8	3.6	15.9
	Average	3.56	4.14	4.34	2.71	4.02	2.28	11.00
Nitrification Rate† (kg NH_4^+-N oxidized/kg MLVSS/day)	Highest	0.055	0.079	0.070	0.062	0.044	0.055	0.146
	Average	0.039	0.045	0.042	0.036	0.037	0.035	0.102
Denitrification Rate (mg NO_3^--N reduced/ℓ reactor volume/hr)	Highest	3.9	4.5	4.6	3.0	5.2	4.6	4.9
	Average	3.23	3.23	3.31	2.65	3.96	2.22	3.73
Denitrification Rate (mg NO_3^--N reduced/g MLSS/hr)	Highest	1.19	1.24	1.14	1.07	1.33	1.71	1.19
	Average	1.00	0.90	0.83	0.95	1.01	0.83	0.90
Denitrification Rate (kg NO_3^--N reduced/kg MLVSS/day)	Highest	0.043	0.049	0.044	0.040	0.048	0.069	0.046
	Average	0.035	0.035	0.032	0.035	0.037	0.034	0.034

* Temperatures are in part estimated in accordance with monthly average temperatures.

+ When computing average rates of denitrification for Flow Sheet IV, only four cells were counted as operative anoxic cells; the remaining two anoxic cells were inoperative due to exhaustion of nitrates.

† Average rate of nitrification refers to the aerobic cells placed upstream of anoxic cells.

denitrification rates observed, between 0.040 and 0.049 mg $NO_2^-+NO_3^-$-N gasified/mg MLVSS/day, are probably more significant than the mean rates computed. This range of highest rates was encountered in all flow sheets profiled.

Several other factors that might affect denitrification rates were not investigated including:

a) Effect of adsorbed carbon, if any

b) Effect of water temperature

c) Effect of CRT

d) Effect of the aeration pattern and sludge management procedures on the share of denitrifying biota in the biomass.

The substantial uniformity of the ENP rates observed nevertheless seems to indicate that the share of the denitrifying biota was little affected by the small changes in aeration pattern implemented, by the change in CRT that occurred in going from Flow Sheet IV to Flow Sheet VI, or by the adsorbed carbon which probably was present under Flow Sheet I.

Evaluation of Flow Sheets I, IV, and VI

Flow Sheet I provided too little denitrification time; too much NO_3^--N was left in the process water at the end of the anoxic period.

Flow Sheet IV provided too little nitrification time, within the framework of the prevailing n growth rate, and too much denitrification time. At the end of the main nitrification period, a considerable amount of NH_4^+-N was left in the process water; the denitrification period contained two idle cells in which practically no reaction occurred.

Flow Sheet VI provided too little denitrification time. A cascade arrangement for nitrification and subsequent denitrification of the ammonia nitrogen released during the first anoxic period might have improved overall N removal performance.

ANALYSIS OF MONTHLY NITROGEN BALANCES: FLOW SHEETS IV, V, AND VI

Rates of Denitrification

The denitrification rates in Table 36, obtained by analysis of the mass balance data in Table 25, appear to be higher on the average than the kinetic rates developed in Table 35 from the nitrogen profiles of Figures 23-29. This might be due to denitrification occurring in the final settler. A 2 mg/ℓ denitrification effect attributable to the final settler would explain much of the difference encountered. Furthermore, it should be noted that the mg/ℓ shown as denitrified in Table 25 were not actually measured, but rather computed by deduction. They are, therefore, subject to considerable error. The rates for Flow Sheet IV have been shown for flow-through-times based on both four and six anoxic compartments. Comparing the denitrification rates for

TABLE 36. MONTHLY RATES OF DENITRIFICATION ABSTRACTED FROM MONTHLY NITROGEN BALANCES FOR FLOW SHEETS IV, V, AND VI

Month	Flow Sheet	Temp. Final Settler Effl. (C)	ML Vol. Fract. (%)	Rates of Denitrification* (mg/g VSS/hr)	(g/g VSS/day)	(mg/g VSS/hr)**	(g/g VSS/day)**	Growth Rate-HM (day^{-1})
Aug.'74	IV	25.5	65	2.09	0.050	3.14	0.075	0.23
Sept.'74	IV	22.5	63	1.60	0.038	2.40	0.058	0.28
Oct.'74	IV	21.5	65	1.22	0.029	1.83	0.044	0.18
Nov.'74	IV	19.3	69	1.07	0.026	1.61	0.039	0.26
Dec.'74+	IV	16.3+	69+	0.89+	0.021+	1.34+	0.032+	0.21+
Jan.'75	IV	16.6	71	1.09	0.026	1.64	0.039	0.16
Feb.'75	IV	15.9	73	1.24	0.030	1.86	0.045	0.22
Mar.'75	IV	18.1	68	1.28	0.031	1.92	0.046	0.19
Apr.'75	V	18.5	62	1.30	0.031	N.AP.	N.AP.	0.16
May'75	VI	23.0	65	1.51	0.036	N.AP.	N.AP.	0.12
June'75	VI	24.5	62	1.82	0.044	N.AP.	N.AP.	0.14
July'75	VI	27.1	59	1.71	0.041	N.AP.	N.AP.	0.11

* Rates of denitrification refer to N removed in ENR.

** Adjusted for two ineffective anoxic cells in the aeration train, Flow Sheet IV only.

HM . . . heterotrophic matrix

+ December data are of little significance because plant was operated as primary plant for part of the month due to breakdown of final settler equipment.

August 1974 with those for January 1975, a theta value (Θ) of 1.08 could be computed, with respect to a reference temperature of 20 C.

The theta symbol is used here in its usual sense: it denotes the base of an exponential factor employed to describe the temperature dependency of a biological reaction rate; a reference temperature has to be selected for which the exponential factor assumes the value one; equation 50 illustrates the use of Θ as here defined; in that equation, Θ has the value 1.06 and the reference temperature selected is 20C. The exponent of Θ is always the difference between reaction temperature and reference temperature. The type of functional dependency between reaction rate and temperature described by equation 50 was stipulated by Arrhenius-Phelps-Streeter. For a given reaction, Θ may be determined mathematically if the reaction rates for two temperatures are known. For such a determination, knowledge of the reaction rate at the reference temperature is not required.

Sludge Production

Actual sludge production for 12 mo is compared in Table 37 with projected sludge production, the latter based on a BOD_5 yield coefficient of 0.9 and a biomass destruction factor of 2.15 mg biomass destroyed/mg N removed via ENR. The results indicate fair agreement, particularly for Flow Sheet IV when the primary settler was not used.

The BOD_5 yield coefficient of 0.9 mg/mg and the biomass destruction factor of 2.15 mg/mg were selected in a trial-and-error procedure to produce good agreement between actual sludge production and sludge production computed on the basis of these two biological coefficients. The BOD_5 yield factor is close to the theoretical yield factor developed in Section 4. The biomass destruction factor of 2.15 mg/mg is considerably smaller than the theoretical value of 2.69 mg/mg developed in Section 4 for ENR. The difference probably reflects the effect of adsorbed carbon nitrate respiration.

Nitrifier Growth Rates Based on Nitrogen Balances

In the preparation of this table, a simplified equation was employed to compute the average PF n. growth rate corresponding to influent and effluent ammonia nitrogen concentrations and environmental conditions, to wit:

$$\mu'_{na} = \frac{\mu_{max}\,(S_0 - S_1)}{(S_0 - S_1) + K_s \ln\left(\dfrac{S_0}{S_1}\right)}$$

This simplified equation disregards the sludge recycle ratio (r). The EPA PDM accepts disregarding r if it is under 1.0, and in this case, it was consistently under 0.5. The pH was assumed to be 7.2. Because of the simplification described, the symbol μ'_{na} was employed instead of the μ_{na} used in equation 27 of Section 4.

The μ'_{na} values determined are juxtaposed in Table 38 to the n. growth rates imposed on the system via sludge wasting of the HM. Significant discrepancies

TABLE 37. MONTHLY ACTUAL VS. PREDICTED VOLATILE SLUDGE PRODUCTION FOR FLOW SHEETS IV, V, AND VI

Month	Flow Sheet	BOD_5	N*	Projected Sludge Production			Actual VSS Produced (mg/ℓ)
		Removed from Sewage in Aeration System		VSS Applicable (1)		Net VSS Production Predicted (mg/ℓ)	
		(mg/ℓ)	(mg/ℓ)	to BOD_5 Removed (mg/ℓ)	to N Removed via ENR (mg/ℓ)		
Aug. '74	IV	178	13.1	160	28	132	118
Sept. '74	IV	155	8.5	140	18	122	122
Oct. '74	IV	190	11.0	171	24	147	136
Nov. '74	IV	177	7.0	159	15	144	140
Dec. '74	IV	164	7.4	148	16	132	130
Jan. '75	IV	194	10.7	175	23	152	161
Feb. '75	IV	201	11.8	181	25	156	173
Mar. '75	IV	192	11.9	173	26	147	144
Apr. '75	V	143	10.6	129	23	106	111
May '75	VI	115	13.2	104	28	76	89
June '75	VI	102	12.4	92	27	65	82
July '75	VI	86	10.7	77	23	54	60

* Via ENR.

(1) Based on 0.9 mg VSS produced/mg BOD_5 removed and on 2.15 mg VSS destroyed/mg N removed via ENR.

TABLE 38. MONTHLY GROWTH RATES ABSTRACTED FROM MONTHLY NITROGEN BALANCES FOR FLOW SHEETS IV, V, AND VI

Month	Temp (C)	S_0 (mg/ℓ)	S_1 (mg/ℓ)	DO (mg/ℓ)	μ'_{na} (day^{-1})	μ_{N1} (day^{-1})	μ_{N2} (day^{-1})	μ_{HM} (day^{-1})	K_s (mg/ℓ)
1974									
Aug.	25.5	10.5	0.4	4	0.68	0.46	0.69	0.23	1.39
Sept.	22.5	9.6	0.6	3	0.53	0.56	0.84	0.28	0.98
Oct.	21.5	10.9	0.3	3	0.48	0.36	0.54	0.18	0.87
Nov.	19.3	12.3	2.0	4	0.48	0.52	0.78	0.26	0.67
Dec.	16.3	12.0	3.2	4	0.37	0.42	0.63	0.21	0.47
1975									
Jan.	16.6	12.8	0.8	4.5	0.38	0.40	0.60	0.16	0.48
Feb.	15.9	14.9	1.2	4.5	0.36	0.44	0.66	0.22	0.45
Mar.	18.1	14.7	2.7	4.0	0.44	0.38	0.57	0.19	0.58
Apr.	18.5	14.7	2.1	4.2	0.42	0.32	0.48	0.16	0.61
May	23.0	16.0	0.5	3.5	0.62	0.24	0.48*	0.12	1.04
June	24.5	14.7	0.2	4.0	0.65	0.28	0.56*	0.14	1.23
July	27.1	11.4	0.1	2.0	0.55	0.22	N.AV.	0.11	1.68

μ'_{na} Nitrifier (Nitrosomonas) growth rate, predicted on the basis of kinetic theory.

μ_{N1} Imposed nitrifier growth rate, assuming that all aerobic cells are growth cells.

μ_{N2} Imposed nitrifier growth rate, assuming that only the aerobic cells of the first aerobic block are growth cells.

μ_{HM} Average growth rate of heterotrophic matrix.

* Nitrification assumed to be complete within the first three aerobic cells.

will be noted.

SOLIDS MANAGEMENT ASPECTS OF FLOW SHEETS IV, V, AND VI

The Relationship Between SVI, RSSS, and MLSS Concentrations. Effect of BOD_5 Load on RSSS Concentrations

The sludge concentration in the space occupied by the settled sludge at the end of a 30-min settled volume test (SS_{SV30}) is related to the SVI as follows:

$$SS_{SV30} = \frac{10^6}{SVI} \tag{40}$$

This equation follows immediately from the definition of the SVI and from geometrical considerations.

SS_{SV30} has been regarded traditionally as approximately equal to the RSSS concentration, provided there is no substantial compaction or dilution of return sludge in the final settler. Such a condition would prevail if the final settler were operated with a rather shallow sludge blanket. SS_{SV30} might be regarded as the maximum safe RSSS concentration that is achievable. In many situations, it will be more practical to operate with a smaller RSSS concentration, i.e., with a higher sludge recycle rate than corresponds to SS_{SV30}.

Using Equation 40, Table 39 was prepared. The table indicates that SS_{SV30} indeed was approximately equal to RSSS concentration, but that the correlation was generally poor. The March 1975 value of the RSSS concentration for instance was 139 percent of the corresponding SS_{SV30} value, while the January 1975 RSSS concentration was only 84 percent of the corresponding SS_{SV30} value.

The RSSS concentration hovered around 12,000 mg/ℓ for the period during which the primary settler was bypassed and near 15,000 mg/ℓ for the period during which primary settling was employed. Table 39 illustrates the beneficial effect of lowered organic load and increased SRT on RSSS concentration and SVI.

Tabel 20 which lists the ratios SV60/SV30 for the 10 mo of August 1974-May 1975 indicates that compaction in the second 30 min of the settled volume test is considerably larger for sludges with a high SVI (in the 90-110 mℓ/g range) - approximately 25 percent - than for sludges with a low SVI (under 70 mℓ/g) - approximately 10 percent.

OPERATING RESULTS: CHLORINE CONTACT CHAMBER/SECOND-PHASE SETTLER COMBINATION

Baffle cages were installed in the combined chlorine contact chamber/second-phase final settler unit (refer to Figure 11) on June 18, 1975. The data in Table 40 for the 5-wk period of June 28-August 1, 1975, indicate an average 33 percent reduction in SS from 5.4 mg/ℓ down to 3.6 mg/ℓ was achieved across the unit.

TABLE 39. MONTHLY-AVERAGE RETURN SLUDGE CHARACTERISTICS FOR FLOW SHEETS IV, V, AND VI

Month	Year	RSSS (mg/ℓ)	SVI (ml/g)	SS_{SV30} (mg/ℓ)	SRT Based on MLSS (days)	Daily BOD_5 Load on MLVSS (kg/kg)
Aug.	1974	11,400	108	9,300	4.4	0.35
Sept.	1974	11,100	91	11,000	3.7	0.37
Oct.	1974	13,300	93	10,800	5.5	0.26
Nov.	1974	12,000	76	13,200	3.8	0.34
Dec.	1974	13,300	64	15,600	4.8	0.28
Jan.	1975	13,100	64	15,600	6.1	0.26
Feb.	1975	11,300	75	13,300	4.5	0.28
Mar.	1975	12,700	110	9,100	5.2	0.27
Apr.	1975	15,000	73	13,700	6.2	0.23
May	1975	14,200	88	11,400	8.1	0.17
June	1975	14,500	68	14,700	7.0	0.19
July	1975	16,600	59	16,900	9.0	0.17

During the 5 wk ending August 1, 1975, the following operating conditions prevailed:

Average daily flow: 629 m^3/day (166,000 gpd)

Average flow rate: 7.28 ℓ/sec (0.255 cfs)

Average flow velocity taken over the 1.525-m x 1.525-m (5-ft x 5-ft) cross section: 0.0031 m/sec (0.0102 ft/sec)

Average flow-through-time for 4.88-m (16-ft) length: 26 min

It is believed the operation of the subject combination unit could be improved by using corrugated sheets instead of the flat Plexiglass sheets as baffles. Using corrugated sheets would drastically cut down the accumulation time of the solids on the baffles. Solids accumulate on the baffles until a certain thickness is reached. At that point, the solids slide off; if the accumulation time is too long, a danger exists that the solids will become septic and float to the top. Chlorination, of course, retards biological action.

The orifice plates had to be cleaned periodically of scraps of papers and bits of rags. In order to prevent such a manual operation, a combined second-phase settler/chlorine contact chamber should be protected by a travelling screen.

SLUDGE CHLORINATION

Data generated during 19 batches of sludge chlorination are summarized in Tables 41, 42, and 43. Characteristics of the underflow from the sludge

TABLE 40. EFFECT OF CHLORINE CONTACT CHAMBER/SECOND-PHASE SETTLER COMBINATION ON SUSPENDED SOLIDS REMOVAL

Date 1975	Influent SS (mg/ℓ)	Effluent SS (mg/ℓ)
June 28	2.0	2.6
29	5.1	3.4
30	7.3	3.1
July 1	5.2	N.AV.
2	3.8	2.7
3	N.AV.	2.7
4	7.9	1.3
5	5.5	4.7
6	6.3	5.1
7	4.5	3.9
8	5.6	4.7
9	6.8	4.5
10	5.0	4.5
11	5.4	3.7
12	5.1	3.5
13	3.8	3.3
14	6.0	4.4
15	4.8	3.5
July 16	3.6	2.7
17	4.2	2.9
18	6.2	2.9
19	3.8	2.7
20	6.3	4.5
21	5.1	3.8
22	5.5	3.9
23	6.7	4.6
24	7.6	5.9
25	7.2	4.5
26	4.9	4.0
27	5.8	3.7
28	11.4	3.7
29	5.4	4.1
30	4.3	3.1
31	4.9	3.8
Aug. 1	6.9	3.9
Ave.	5.4	3.6

dewatering beds for nine of those batches are presented in Table 44.

The batch size varied somewhat according to the solids concentration of the sludge treated and also according to the sludge storage space available. A typical batch consisted of 34 m^3 (9000 gal) applied to 111 m^2 (1200 ft^2) of sand bed, resulting in an average sludge slurry dose of 0.31 m (1 ft). However, such a depth was never attained because drainage occurred immediately upon application. At most, approximately 15 cm (6 in.) of chlorinated sludge slurry were seen standing on the bed. Usually, all visible liquid had drained away by the morning following application.

The slurry was treated at a rate of approximately 2.5 ℓ/sec (40 gpm). The weighted average of the chlorine dosage was 830 mg/ℓ or 10 percent of the dry weight of the sludge chlorinated. The chlorine dosage was adjusted to produce a pH of 2.3 to 2.8 in the chlorinated slurry. The pH of the unchlorinated slurry indicates that considerable nitrification had occurred on some batches before treatment.

During chlorination, samples of unchlorinated and chlorinated sludge slurry were withdrawn and subjected to the tests reflected in Tables 41, 42, and 43. Glass fiber filters were used to determine some of the parameters in Table 44.

There was only a slight reduction in SS due to chlorination, approximately 5 percent. It should be noted in this connection that solids reduction due to acidification with H_2SO_4 to pH 2 was found to be between 50 and 60 percent in unrelated laboratory tests on the waste sludge of this plant.

There was an approximate 2 percent increase in the VSS concentration, probably attributable to experimental error.

The increase in filterability, 49 percent on the average, was of course the most important change in sludge characteristics brought about by sludge chlorination. The increase of filterability was greater for sludges of a low initial filterability. The weighted average filterability was increased from 52 to 101 mℓ/30 sec. The mechanisms of this increase remain to be explained.

The changes in nutrient concentration were somewhat erratic but on the average negligible, i.e., under 3 mg/ℓ.

The chlorine residual in the chlorinated sludge was approximately 150 mg/ℓ. A significant increase in the TOC of the sludge supernatant was observed from a range of 11-150 mg/ℓ to a range of 46-188 mg/ℓ.

The sludge dewatering beds underflow did not impose any significant load on the A.S. system. The following constituent ranges were determined:

SS	9-76 mg/ℓ
VSS	2-43 mg/ℓ
pH	5.2-6.9
Chlorine Residual	0.1-5.3 mg/ℓ
TOC	39-60 mg/ℓ
COD	142-241 mg/ℓ

Alkalinity	42-130 mg/ℓ as $CaCO_3$
NH_4^+-N	6-36 mg/ℓ
$NO_2^-+NO_3^-$-N	1-17 mg/ℓ
Dissolved Orthophosphorus	0.4-1.8 mg/ℓ

TABLE 41. CHLORINE DOSAGES USED DURING SLUDGE CHLORINATION EXPERIMENTS

Date	Batch	Volume of Sludge Clorinated		Cl_2 Used		SS of Sludge Before Treatment	Dry Sludge Treated		Chlorine Dosage	
1975	No.	(m^3)	(gal)	(kg)	(lb)	(mg/ℓ)	(kg)	(lb)	(mg/ℓ)	% of dry solids
7/8	1	31.8	(8,400)	22	(48)	9,200	293	(645)	685	7
7/9	2	31.0	(8,200)	27	(60)	7,400	230	(506)	877	12
7/10	3	31.4	(8,300)	25	(54)	10,100	318	(699)	780	8
7/18	4	22.4	(6,000)	24	(52)	8,700	198	(435)	1039	12
7/22	5	33.7	(8,900)	36	(78)	10,700	361	(794)	1051	10
7/28	6	33.7	(8,900)	33	(72)	12,700	429	(943)	970	8
7/29	7	37.8	(10,000)	21	(46)	8,700	330	(726)	552	6
7/31	8	32.6	(8,600)	24	(54)	13,400	437	(961)	753	6
8/6	9	23.5	(6,200)	25	(55)	9,600	225	(496)	1064	11
8/12	10	31.0	(8,200)	36	(80)	9,700	301	(663)	1170	12
8/13	11	32.6	(8,600)	18	(40)	9,800	320	(703)	558	6
8/26	12	68.6	(18,100)	59	(129)	7,300	501	(1102)	855	12
8/29	13	22.7	(6,000)	24	(52)	7,400	168	(370)	1039	14
9/3	14	51.9	(13,700)	38	(84)	6,300	327	(720)	735	12
9/8	15	37.8	(10,000)	35	(76)	7,100	269	(592)	911	13
9/9	16	45.0	(11,900)	30	(66)	5,400	244	(536)	665	12
9/11	17	45.0	(11,900)	24	(54)	5,600	253	(556)	544	10
10/6	18	22.7	(6,000)	24	(54)	7,200	164	(360)	1079	15
10/7	19	42.4	(11,200)	39	(87)	4,200	178	(392)	931	22
Total		678	(179,100)	564	(1241)		5,546	(12,199)	831*	10*

Batch # 18: Primary sludge; all other batches were a mixture of primary sludge and EAS.
* Weighted mean.

TABLE 42. EFFECT OF SLUDGE CHLORINATION ON SOLIDS CONCENTRATION, FILTERABILITY, AND pH

Date 1975	Batch No.	SS Before (mg/ℓ)	SS After (mg/ℓ)	SS Change (%)	VSS Before (mg/ℓ)	VSS After (mg/ℓ)	VSS Change (%)	Filterability Before (ml/30 sec)	Filterability After (ml/30 sec)	Filterability Change (%)	pH Before	pH After
7/8	1	9,190	8,680	-6	4,950	5,020	+1	122	182	+49	6.45	2.6
7/9	2	7,400	7,240	-2	3,780	3,680	-3	124	172	+40	6.3	2.65
7/10	3	10,100	9,540	-6	5,680	5,580	-2	20	97	+385	7.0	2.5
7/18	4	8,730	8,070	-8	5,100	5,030	-1	158	230	+46	6.7	2.5
7/22	5	10,690	10,420	-3	6,190	6,470	+5	16	69	+331	6.75	2.7
7/28	6	12,705	11,765	-7	7,480	7,390	-1	12	67	+458	7.1	2.4
7/29	7	8,680	8,020	-8	4,100	4,440	+8	120	162	+35	6.0	2.5
7/31	8	13,440	12,620	-6	7,040	7,420	+5	14	63	+350	6.4	3.0
8/6	9	9,590	9,220	-4	5,910	6,010	+2	10	64	+540	6.6	3.1
8/12	10	9,680	9,720	-	5,640	6,080	+8	34	76	+124	6.7	2.75
8/13	11	9,780	9,460	-3	5,550	5,600	+9	68	99	+46	6.0	3.0
8/26	12	7,300	6,920	-5	4,860	4,820	-1	16	67	+319	6.5	2.6
8/29	13	7,360	5,920	-20	4,720	4,130	-13	88	144	+64	5.85	2.4
9/3	14	6,260	6,710	-7	4,100	4,890	+19	84	152	+80	6.9	3.2
9/8	15	7,050	6,330	-10	4,220	4,030	-5	50	84	+68	6.8	2.4
9/9	16	5,440	4,950	-9	3,320	3,340	-	51	82	+61	6.9	2.3
9/11	17	5,600	5,430	-3	3,460	3,560	+3	30	77	+157	6.9	2.5
10/6	18	7,220	6,940	-4	4,580	4,820	+5	10	83	+730	7.2	2.25
10/7	19	4,220	4,020	-5	2,780	2,920	-5	22	24	+10	7.4	2.7
Weighted Mean		8,180	7,760	-5	4,790	4,900	+2	52	101	+49		

TABLE 43. EFFECT OF SLUDGE CHLORINATION ON SLURRY FILTRATE QUALITY

Batch No.	TOC			NH_4^+-N			$NO_2^-+NO_3^-$-N			Dissolved Orthophosphorus			Chlorine Residual
	Before (mg/ℓ)	After (mg/ℓ)	Change (%)	Before (mg/ℓ)	After (mg/ℓ)	Change (mg/ℓ)	Before (mg/ℓ)	After (mg/ℓ)	Change (mg/ℓ)	Before (mg/ℓ)	After (mg/ℓ)	Change (mg/ℓ)	After (mg/ℓ)
1	17.5	63.5	+363	1.3	2.3	+ 1.0	42.0	42.0	0	0.5	2.4	+1.9	133
2				0.6	1.2	+ 0.6	74.0	71.5	- 2.5	1.8	8.8	+7.0	102
3				8.9	9.2	+ 0.3	0.2	1.0	+ 0.8	2.1	1.6	-0.5	111
4				1.0	1.7	+ 0.7	35.5	34.5	- 1.0	1.9	6.1	+4.2	233
5				9.5	13.6	+ 4.1	37.5	31.2	- 6.3	2.2	2.2	-	142
6	115	184	+183	20.5	19.6	- 0.9	5.7	0.4	- 4.3	1.6	2.0	+0.4	164
7	11	71	+545	0.5	2.2	+ 1.7	59.0	56.0	- 3.0	0.6	2.6	+2.0	191
8	150	188	+253	22.0	22.0	0	62.5	83.5	+21.0	5.1	3.0	-2.1	121
9				72.6	61.2	-11.1	0.3	0.3	-	4.7	1.3	-3.4	22
10				1.3	3.6	- 2.3	5.1	19.9	+14.8	0.8	0.8	-	89
11				2.8	28.5	+25.7	3.8	15.0	+11.2	0.4	1.6	+1.2	150
12				1.7	0.2	- 1.5	7.6	4.0	- 3.6	0.92	1.5	+0.6	126
13	20.5	46	+124	1.6	2.4	+ 0.8	2.7	0.8	- 1.9	0.20	3.15	+3.0	244
14	24	61.5	+156	8.8	11.7	+ 2.9	0.8	6.9	+ 6.1	0.31	1.40	+1.1	121
15	66	74.5	+ 13	0.1	9.1	+ 9.0	6.5	4.3	- 2.2	4.53	5.65	+1.1	171
16	20	49	+145	0.4	0.5	+ 0.1	0.5	1.5	+ 1.0	0.33	4.70	+4.4	197
17	39	50	+ 28	0.2	0.1	- 0.1	3.8	3.7	- 0.1	0.61	1.95	+1.3	74
18				0.1	0.3	+ 0.2	1.7	0.9	- 0.8	3.32	5.50	+2.2	315
19				0.2	0.2	-	8.7	0.6	- 8.1	0.92	3.96	+3.0	168
Weighted Mean				6.7	8.7	+ 2.0	17.6	18.4	+ 0.8	1.6	3.0	+1.4	146

TABLE 44. QUALITY OF SLUDGE DEWATERING BED UNDERFLOW DURING SLUDGE CHLORINATION EXPERIMENTS

			Constituents of Bed Filtrate**									
Batch No.	Type of Bed*	Chlorine Dosage (mg/ℓ)	SS (mg/ℓ)	VSS (mg/ℓ)	pH	Chlorine Residual (mg/ℓ)	TOC (mg/ℓ)	COD (mg/ℓ)	Total Alk. (mg/ℓ as $CaCO_3$)	NH_4^+-N (mg/ℓ)	NO_2^-+ NO_3^--N (mg/ℓ)	Dissolved Orthophosphorus (mg/ℓ)
11	CTF	558	21	13	6.6	2.7	N.AV.	193	78	8	5	1.4
12	CTF	855	76	24	6.2	0.7	N.AV.	208	115	12	15	1.2
13	SSB	1039	29	12	5.2	0.1	43	147	42	36	11	1.8
14	CTF	735	14	N.AV.	6.3	2.7	39	110	130	11	1	0.4
15	CTF	911	40	19	6.7	0.1	44	241	90	24	17	1.3
16	SSB	665	9	2	5.8	5.3	51	142	87	12	14	0.5
17	CTF	544	64	43	6.7	0.1	60	178	73	15	12	1.0
18	CTF	1079	25	22	6.3	0.1	N.AV.	199	52	21	8	1.0
19	CTF	931	55	27	6.9	2.7	N.AV.	211	127	6	15	1.3

* CTF = Converted trickling filter.
SSB = Standard sand drying bed.

** Grab samples.

SECTION 9

DESIGN CONSIDERATIONS FOR ACTIVATED SLUDGE SYSTEMS EMPLOYING COMPARTMENTALIZED AERATION TANKS

BASIC DESIGN ALTERNATIVES

The reactor should be designed as a compartmentalized tank to produce a near PF hydraulic regime and to allow for flexibility in the aeration pattern. All biological reactions deviate from the zero order reaction mode in the area of low substrate or hydrogen acceptor ($NO_2^-+NO_3^--N$) concentration, and such low concentrations are often prescribed under today's more stringent effluent requirements. The PF hydraulic regime is obviously superior to the CM hydraulic regime for all reactions of an order greater than zero. Aeration pattern flexibility is important to achieve full reactor utilization in all seasons.

The block arrangement of aerobic-anoxic cells is one of the two basic patterns of aeration cells that seems to be available, as illustrated previously in Figure 3. The other arrangement is the alternating zones approach. Flexible mechanical design will allow the use of either approach in the same reactor. The alternating zones pattern was not tested for a sufficient period of time on this project to adequately assess its potential. However, the alternating zones approach, if used in a reactor of generous overall design, might make it unnecessary to adjust the aeration pattern to seasonal changes of operating conditions.

A recommended cell size is 9.1 m x 9.1 m x 9.1 m SWD (30 ft x 30 ft x 30 ft), equivalent to a cell volume of 754 m^3 (202,000 gal). For gentle mixing, a power input of 10 W/m^3 is required according to Bradley (23). The inlet and outlet ports have to be near the bottom because the solids are not uniformly dispersed throughout the tank cells. The size of these ports should be carefully computed so as not to cause undue differences in the SWD of the cells. DO control equipment should be provided for any cell where aeration equipment is installed. An excessive DO concentration, over 3 mg/ℓ, will impair the efficiency of the following anoxic cell.

A large sludge recycle rate will depress the substrate gradient in a compartmentalized reactor; this is kinetically disadvantageous because the average substrate concentration in a PF reactor determines its removal rate. On the other hand, a high recycle rate often makes management of the solids in the final settler easier.

Exclusive of the aeration system itself, a decision must be made whether to provide either primary settling or flow equalization or both. To cope with load fluctuations, the designer has the choice, in many cases, between a larger aeration tank and flow equalization. In a PF aerobic-anoxic reactor, extension of flow-through time up to a certain point will not cause the release of nutrients nor deterioration of the effluent via deflocculation, assuming that P is controlled by metal salt addition; it will, however, result in decreasing sludge production. The designer should also be cognizant that flow equalization imposes a certain measure of complexity on the operation and maintenance of a plant.

It is, therefore, recommended that increasing the reactor size be given priority consideration unless there are other factors beside the secondary system that can only be resolved by flow equalization. If flow equalization is not included in the design, a deep secondary settler must be provided to accommodate diurnal expansion of the sludge blanket. Similar considerations apply to the design of the primary settler. Another point to consider is that systems not equipped with primary settling will possess greater N removal capacity.

In both aeration patterns, the last cell prior to the final settler should be an aerobic cell to nitrify any ammonia nitrogen released in the preceding anoxic cells and to purge any nitrogen gas clinging to the floc. If such an aerobic cell is not provided, settling difficulties may occur.

REACTOR SIZING EQUATIONS

In this discussion, reactor volumes are expressed in terms of sewage detention times. The dimension is day. SDT, SDT_n, SDT_d, and SDT_p refer to the entire reactor, nitrification reactor, denitrification reactor, and purge reactor, respectively. By multiplying sewage detention time values by the daily plant flow, the various reactor volumes are obtained. If the plant flow is expressed as m^3/day, plant capacity is calculated as m^3, etc. The purge reactor serves to nitrify NH_4^+-N released during SDT_d. There is no need to determine a reactor size requirement for aerobic respiration/synthesis. That reaction occurs during nitrification. Many of the kinetic coefficients appearing in the design equations have been insufficiently researched so far. The design equations suggested should, therefore, be used only to determine design parameters for a pilot plant. The design of a municipal plant should be preceded by fairly large-scale pilot plant studies, preferably on the site of the proposed facility. Values of kinetic coefficients suggested on the basis of this project are listed in the following subsection which presents a design example.

Consistent with Section 4 and the EPA PDM, no decay coefficient for nitrifiers is used in the reactor design equations presented in the following paragraphs; however, a decay coefficient (b) for the heterotrophic biomass is employed in the conventional way to account for biomass destruction through auto-oxidation in aerobic cells.

Equations 41 to 46 are suggested for the computation of SDT_n, SDT_d, SDT_p, and SDT, to be used in the sequence listed. Except as regards SDT_p, the design equations are the same for both the block and the alternating zones approach. Under the block approach, the purge reactor has to nitrify all the NH_4^--N released during SDT_d. Under the alternating zones approach, only the NH_4^+-N released in the last anoxic zone has to be nitrified during SDT_p.

$$SDT_n = \frac{SF_n \left[\Delta SC\ Y - b_d\ (N_0 - N_1)\right]}{X \left[\mu_{nr}\ (1 - n\ b_d) + b\ SF_n\right]} \tag{41}$$

$$\Delta X = \frac{SDT_n\ \mu_{nr}\ X}{SF_n} \tag{42}$$

$$N_d = N_0 - N_1 - \Delta X\ n \tag{43}$$

$$SDT_d = \frac{N_d\ SF_d}{k_d\ X} \tag{44}$$

$$SDT_p = N_d\ b_d\ n \left[\frac{\mu_{nr}\ SDT_n}{\mu_{nsl}\ (S_0 - S_1)\ SF_n}\right] \tag{45}$$

$$SDT = SDT_n + SDT_d + SDT_p \tag{46}$$

where:

SF_n = safety factor, nitrification reactor

ΔSC = BOD_5 removed (during SDT_n) per liter of sewage treated, mg/ℓ

b_d = biomass (VSS) destroyed per mg of NO_3^--N reduced in ENR, mg/mg

N_0 = biodegradable nitrogen in the influent, mg/ℓ

N_1 = biodegradable nitrogen in the effluent, mg/ℓ

n = decimal nitrogen fraction in the wasted biomass

b = microorganism decay factor in EOR, day^{-1}

N_d = NO_3^--N reduced to gaseous nitrogen during SDT_d per liter of sewage treated, mg/ℓ

SF_d = safety factor, denitrification reactor (ENR)

k_d = specific ENR reaction rate (mg NO_3^--N reduced/mg MLVSS/day), day^{-1}

S_0 = nitrification reactor influent NH_4^+-N concentration, mg/ℓ

S_1 = nitrification reactor effluent NH_4^+-N concentration, mg/ℓ

X = mixed liquor volatile suspended solids concentration (biomass concentration in the ML), mg/ℓ

ΔX = Net biomass production (mg VSS synthesized less mg VSS destroyed in EOR and ENR) per liter of sewage treated.

Y = yield factor, BOD_5 removal (mg VSS produced/mg BOD_5 removed at zero cell residence time), mg/mg

μ_{nr} = specific n. growth rate to be imposed on a reactor to satisfy given effluent requirements, day^{-1}

μ_{ns1} = specific n. growth rate corresponding to assumed NH_4^+-N concentration in effluent of CM purge reactor, day^{-1}.

To make the mathematics manageable, the small amount of biomass destruction that occurs in the purge reactor was neglected in the development of the above equations. The effect of this omission is estimated in the following subsection.

Remarks Regarding Kinetic Coefficients and SF_d

Regarding b_d --

The ENR biomass destruction factor is difficult to measure directly in pilot plant experiments. However, b_d may be determined indirectly by using the following equation, which follows readily from the definition of b_d:

$$b_d = \frac{\text{mg } NH_4^+\text{-N released/mg } NO_3^-\text{-N reduced}}{\text{nitrogen fraction in wasted biomass}} \qquad (47)$$

Regarding μ_{nr} --

The calculation of μ_{nr} was discussed at some length in Section 4. If the nitrification reactor is a PF reactor, S_0, S_1, and r must be given or assumed. If the nitrification reactor is a CM reactor, only S_1 must be known.

Regarding $(1 - n\ b_d)$ --

This expression indicates the stoichiometric efficiency (E_{ENR}) of ENR, which will vary in accordance with the adsorbed carbon utilized.

$$E_{ENR} = \frac{\text{N removed in ENR}}{NO_3^-\text{-N reduced in ENR}} = 1 - n\ b_d \qquad (48)$$

Regarding b --

A wide range of values for b, from 0.048 - 0.2 day^{-1}, is listed in the literature (3) (13) (24) (25). The dependency of b on temperature is not discussed in connection with these data listings. This might be due to the fact that b is usually measured as it manifests itself in the form of net sludge production Y' per Equation 49, to wit:

$$Y' = Y - b\left(\frac{X\ SDT_n}{\Delta SC}\right) \qquad (49)$$

If Y is also dependent on temperature, the temperature dependency of b might not become evident. Some researchers (26) have determined a theta (Θ) of 1.07 for b, with respect to the Streeter-Phelps version of the Arrhenius equation.

Regarding k_d --

The product k_d b_d might be called the anoxic counterpiece of b. The following equation is suggested as describing the temperature dependency of k_d:

$$k_d = k_{d20}\ 1.06^{(T-20)} \tag{50}$$

In other words, a Θ of 1.06 is suggested.

Regarding SF_d --

It seems to make little sense to use a large SF_d. Heterotrophs are not as easily inhibited as nitrifiers. The anoxic reactor is somewhat shielded from upsetting substances by the preceding aerobic reactor. Also, it is important to keep the anoxic residence time as short as possible so as not to damage the sensitive nitrifiers.

Choice of Design Parameter Values

An analysis of Equations 41 to 46 indicates that X is the only design variable which may be selected by the designer, once he has chosen the effluent NH_4^+-N concentration of the nitrification reactor (S_1) and r if a PF reactor is involved. While the choice of X affects SDT_n and SDT_d, the ratio SDT_n/SDT_d and the values of ΔX and N_d are not affected by X, once r and S_1 have been chosen.

Derivation of Equations

Equation 41 --

This equation for SDT_n was obtained by combining the following equations:

$$SDT_n = \frac{SF_n\ \Delta X}{\mu_{nr}\ X} \tag{51}$$

$$N_0 = N_1 + N_d + \Delta X\ n \tag{52}$$

$$\Delta X = \Delta SC\ Y' - N_d\ b_d \tag{53}$$

$$Y' = Y - b\left(\frac{X\ SDT_n}{\Delta SC}\right) \tag{49}$$

where: Y' = net yield coefficient in nitrification reactor (VSS produced/mg BOD_5 removed taking biomass destruction due to EOR into consideration), mg/mg.

Equation 51 follows readily from Equations 31 and 34. Equation 52 is a material balance equation with respect to nitrogen, taken over SDT_n and SDT_d. Equation 53 is a material balance equation for biomass, taken over the same two reactors. Equation 49 is the well known equation for sludge production in an aerobic process, accounting for biomass generation by synthesis and for biomass destruction by auto-oxidation.

Equation 44 --

This equation expresses the position that ENR can be regarded as a zero order reaction both with respect to the available NO_3^--N concentration and the amount of biomass available for bio-oxidation.

Equation 45 --

This equation for the calculation of SDT_p was obtained by combining the following two equations:

$$\frac{N_d \; b_d \; n}{SDT_p} = \left(\frac{\mu_{nsl}}{Y_n}\right) \dot{N} \tag{54}$$

$$\dot{N} = \frac{Y_n \; (S_0 - S_1) \; SF_n}{\mu_{nr} \; SDT_n} \tag{55}$$

Equation 54 is a material balance equation (nitrogen) for the CM purge reactor. Equation 55 defines the quantity of $\dot{N}$ developed during SDT_n. Equation 55 follows readily from Equations 20 and 34 of Section 4. It should be noted that the growth rate in the nitrification reactor will not be μ_{nr} but μ_{nr}/SF_n. Also, S_1 is the effluent NH_4^+-N concentration of the nitrification reactor taking the effect of SF_n into consideration.

PROCESS DESIGN EXAMPLE

Data given or prescribed:

BOD_5 to be removed: $\Delta SC = 150$ mg/ℓ

Influent N concentration: $N_0 = 25$ mg/ℓ (biodegradable TKN)

Effluent N concentration: $N_1 = 2$ mg/ℓ (biodegradable total N)

Influent NH_4^+-N concentration: $S_0 = 15$ mg/ℓ

Temperature: T = 10 C (Winter) T = 20 C (Summer)

ML pH = 7.2

ML DO = 3 mg/ℓ

Design variables selected:

MLVSS: $X = 3000$ mg/ℓ

Sludge recycle ratio: $r = 0.5$

Hydraulic regime: PF

Safety factors: $SF_n = 1.5$ $SF_d = 1.25$

Effluent NH_4^+-N concentration, nitrification reactor: $S_1 = 0.5$ mg/ℓ

Kinetic coefficients assumed:

$b = 0.06\ \text{day}^{-1}$	$k_{d20} = 0.040\ \text{day}^{-1}$
$Y = 1.00$	$b_d = 2.5$
$n = 0.08$	

STEP I. For nitrification and ENR, compute the values of the kinetic coefficients that are dependent on environmental conditions and hydraulic regime using Equations 16, 17, 18, 27, and 50.

10 C Values

$$K_s = 10^{[0.051(10) - 1.158]} = 0.225\ \text{mg}/\ell$$

$$\mu_{max} = 0.47\ e^{[0.098(10-15)]}\left(\frac{3}{1.3 + 3}\right) = 0.201\ \text{day}^{-1}$$

$$\mu_{nsl}(CM) = \frac{0.201(0.5)}{0.225 + 0.5} = 0.139\ \text{day}^{-1}$$

$$\mu_{nr}(PF) = \frac{0.201(15 - 0.5)}{15 - 0.5 + (1 + 0.5)\ 0.225\ \ln\left[\frac{15 + 0.5(0.5)}{(1 + 0.5)\ 0.5}\right]} = 0.188\ \text{day}^{-1}$$

$$k_d = 0.040(1.06)^{(10-20)} = 0.022\ \text{day}^{-1}$$

20 C Values

$K_s = 0.728$ mg/ℓ; $\mu_{max} = 0.535$ mg/ℓ

$\mu_{nsl}(CM) = 0.218\ \text{day}^{-1}$; $\mu_{nr}(PF) = 0.436\ \text{day}^{-1}$

$k_d = 0.040\ \text{day}^{-1}$

STEP II. Compute the size of the nitrification reactor using Equation 41 (10 C):

$$SDT_n = \frac{1.5\ [150(1) - 2.5(25-2)]}{3000\ \{0.188\ [1 - 0.08(2.5)] + 0.06(1.5)\}} = 0.192\ \text{day} = 4.62\ \text{hr}$$

STEP III. Compute VSS production per liter of sewage treated using Equation 42 (10 C):

$$\Delta X = \frac{0.192\ (0.188)\ (3000)}{1.5} = 72.2\ \text{mg}/\ell$$

This amounts to 0.48 mg VSS/mg BOD_5 removed, a value well within the range of A.S. process experience.

STEP IV. Compute N to be gasified in ENR per liter of sewage treated using Equation 43 (10 C):

$$N_d = 25 - 2 - 72.2(0.08) = 17.2\ \text{mg}/\ell$$

STEP V. Compute the size of the denitrification reactor using Equation 44 (10 C):

$$SDT_d = \frac{17.2\ (1.25)}{0.022\ (3000)} = 0.326\ \text{day or}\ 7.82\ \text{hr}$$

To avoid possible damage to the nitrifiers, the anoxic reactor should be divided into two parts, separated by a short CM aerobic reactor to nitrify the NH_4^+-N released in the first anoxic tank. Assuming that the first anoxic reactor has a detention time of 4.0 hr = 0.167 day, this division of the ENR process may be expressed as follows:

$$SDT_{d2} = 0.326 - SDT_{d1} = 0.326 - 0.167 = 0.159\ \text{day} = 3.82\ \text{hr}$$

The reactor space added by SF_d will not affect the reaction under design conditions. When reversing the computation and calculating the effect of SDT_{d2} on NO_3^--N concentration and biomass destruction, SF_d will have to be disregarded.

STEP VI. Compute the size of the purge reactor using Equation 45 (10 C):

$$SDT_p = 17.2\ (2.5)\ (0.08) \left[\frac{0.188\ (0.192)}{0.139\ (15-0)\ (1.5)}\right] = 0.040\ \text{day} = 0.95\ \text{hr}$$

The purge reactor should be divided into two compartments of approximately 0.02-day detention time each. While the prescribed or assumed effluent NH_4^+-N concentration (S_1) is 0.5 mg/ℓ, in the above computation, S_1 was assumed to be 0 due to the effect of SF_n. The division of the purge reactor into two compartments may be described by: $SDT_p = SDT_{p1} + SDT_{p2}$. The computation of SDT_p is based on the amount of NH_4^+-N to be oxidized (S_0), on the assumed effluent NH_4^+-N concentration (S_1) which controls μ_{ns1}, and on the assumed N_d concentration. If any of these three factors is changed, SDT_p will have to be recomputed.

STEP VII. Tabulate the reactor detention times in sequence:

Nitrification reactor: SDT_n = 0.192 day = 4.62 hr
First denitrification reactor: SDT_{d1} = 0.167 day = 4.00 hr
First purge reactor: SDT_{P1} = 0.020 day = 0.48 hr
Second denitrification reactor: SDT_{d2} = 0.159 day = 3.82 hr
Second purge reactor: SDT_{P2} = 0.020 day = 0.48 hr

TOTAL: SDT = 0.558 day or 13.39 hr

STEP VIII. To check the results of computation Steps II to VI, trace the fate of the nitrogen forms and the newly formed biomass through the various zones of the reactor, as illustrated by Table 45.

NOTES (Table 45):

(1) For the synthesis reaction which produces 150.00 mg/ℓ new biomass, 12.00 mg/ℓ of N must be removed from the process water. The reactor influent contains 10.00 mg/ℓ of organic N and 15.00 mg/ℓ of ammonia N. Obviously, one has to make an assumption as to the shares of these two forms in the 12.00 mg/ℓ of N that are metabolized in the synthesis reaction. It has been assumed here that all the influent organic N is metabolized (10.00 mg/ℓ) and that the influent ammonia nitrogen contributes only 2.00 mg/ℓ to the total N amount needed for synthesis. It is generally accepted that all particulate matter in a reactor's influent is very quickly adsorbed by the biomass in an aerobic process. The nitrogen profiles of this report indicate that a portion of the dissolved organic influent nitrogen is removed quickly from the process water, while the balance of this nitrogen form remains unchanged through the remainder of the process; apparently this balance can be regarded non-biodegradable. In this design example, the presence of non-biodegradable N compounds has been disregarded.

(2) Biomass destruction: $b \times SDT_n$

(3) A small amount of organic N is first metabolized, then later released as NH_4^+-N. Due to SF_n, S_1 is practically zero.

(4) 11.02 = 0.022 (3000) (0.167)

(5) 1.69 = 2.19 - 0.50. The corresponding size of the purge reactor would be 0.020 day.

(6) The gasification capacity of the second anoxic reaction is 0.022 (3000) (0.159) = 10.49 mg/ℓ. Due to SF_d, only a part of this capacity is utilized under design conditions.

(7) 1.30 = 1.80 - 0.50. The corresponding size of the purge reactor would be:

$$1.30 \left[\frac{0.188\ (0.192)}{0.139\ (15 - 0)\ (1.5)}\right] = 0.015 \text{ day} = 0.36 \text{ hr.}$$

TABLE 45. FORMS OF NITROGEN AND MASS OF NEWLY FORMED BIOMASS AS AFFECTED BY RESIDENCE IN THE FIVE ZONES OF THE REACTOR OF THE PROCESS DESIGN EXAMPLE#

	Organic N in Substrate (mg/ℓ)*	NH_4^+-N in Substrate (mg/ℓ)*	NO_3^--N in Substrate (mg/ℓ)*	N in New Biomass (mg/ℓ)*	Gasified N (mg/ℓ)*	Newly Formed Biomass (mg/ℓ)*
Influent	10.00	15.00				
Nitrification Reactor: Synthesis (1)	-10.00	-2.00		+12.00		+150.00
Nitrification Reactor: Biomass destruction (2)		+2.76		-2.76		-34.56
Nitrification Reactor: Nitrification (3)		-15.76	+15.76			
Balance	0	0	15.76	9.24		115.44
First Anoxic Reactor: Gasification of N (4)			-11.02		+11.02	
First Anoxic Reactor: Biomass destruction		+2.19		-2.19		-27.40
Balance		2.19	4.74	7.05	11.02	88.04
First Purge Reactor: Nitrification (5)		-1.69	+1.69			
Balance		0.50	6.43	7.05	11.02	88.04
Second Anoxic Reactor: Gasification of N (6)			-6.43		+6.43	
Second Anoxic Reactor: Biomass destruction		+1.30		-1.30		-16.23
Balance		1.80	0	5.75	17.45	71.81
Second Purge Reactor: Nitrification (7)		-1.30	+1.30			
Balance or Total (8)		0.50	1.30	5.75	17.45	71.81

* mg/ℓ of sewage treated.

# Biomas destruction in purge reactors neglected.

(8) Due to the effect of SF_n, the overall nitrogen removal result is slightly better than required: N_1 = 1.8 mg/ℓ (computed) vs. N_1 = 2.0 mg/ℓ (prescribed). However, the effect of neglecting biomass destruction in the purge reactors remains to be investigated. ΔX and N_d in Table 45 differ by minute amounts from the values computed under Steps III and IV, respectively, again due to the effect of SF_n.

STEP IX. Compute the aerobic and anoxic reactor sizes for the same treatment situation, but for 20 C.

Using the procedures of Steps II to VI, the following values were obtained. For comparison, the 10 C values are also listed in parentheses.

SDT_n = 0.105 day (0.192)
SDT_d = 0.164 day (0.326)
SDT_p = 0.029 day (0.040)
TOTAL 0.298 day (0.558)

ΔX = 91.6 mg/ℓ (72.2)
N_d = 15.7 mg/ℓ (17.2)
Biomass lost in EOR = 18.9 mg/ℓ (34.56)
$\frac{\Delta X}{\Delta SC}$ = 0.61 (0.48)

The required SDT is 87 percent longer at 10 C than at 20 C, or in other words, only 53 percent of the reactor volume designed for 10 C is needed for treatment at 20 C. The excess reactor capacity available during higher temperature periods could be used for biomass destruction. In other words, much greater safety factors are inherent to summertime operation. Reduction of X or taking some reactor space out of service are two possible ways of adjusting operations to the higher biological reaction rates that prevail during the warmer season.

In the preceding discussion, biomass destruction in the purge reactor was neglected. In the following paragraph, the effect of this omission on overall N removal is estimated.

In Step VI, the size of the purge reactor was computed to be 0.040 day at 10 C. This SDT_p would cause a biomass destruction = b (X) (SDT_p) = 0.06 (3000) (0.040) = 7.2 mg/ℓ sewage treated, thereby releasing (biomass destroyed) (n) = 7.2 (0.08) = 0.58 mg/ℓ NH_4^+-N. Half of this amount would be released in the first section of the purge reactor and would be nitrified instantaneously. In the following second denitrification reactor, these 0.29 mg/ℓ NO_3^--N would be gasified as this reactor zone has excess N reduction capacity due to its safety factor (SF_d). This gasification is accompanied by the release of b_d (NO_3^--N) (n) = 2.5 (0.29) (0.08) = 0.06 mg/ℓ NH_4^+-N resulting from biomass destruction. The total increase in effluent N would, therefore, be equal to the NH_4^+-N released and nitrified from biomass destruction in the second purge reactor plus the NH_4^+-N released in the preceding second denitrification reactor (which would also, of course, be nitrified in the second purge reactor) = 0.29 + 0.06 = 0.35 mg/ℓ, an insignificant amount partially offset by the excess removal effect discussed under Step VIII.

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APPENDICES

Page

A. Derivation of Equation 27 142

B. Data from Aeration Train Profile Samplings 144

  Table B-1. Aeration Train Profile Data for Flow Sheet I (11/14/73) 144

  Table B-2. Aeration Train Profile Data for Flow Sheet I (11/27/73) 145

  Table B-3. Aeration Train Profile Data for Flow Sheet I (12/5/73) 146

  Table B-4. Aeration Train Profile Data for Flow Sheet IV (9/18/74) 147

  Table B-5. Aeration Train Profile Data for Flow Sheet IV (10/10/74) 148

  Table B-6. Aeration Train Profile Data for Flow Sheet IV (11/14/74) 149

  Table B-7. Aeration Train Profile Data for Flow Sheet VI (6/12/75) 150

APPENDIX A

DERIVATION OF EQUATION 27

Equation 27, to wit:

$$\mu_{nr} = \frac{\mu_{max} (S_0 - S_1)}{S_0 - S_1 + (1 + r) K_s \ln \left[\frac{S_0 + r S_1}{(1 + r) S_1}\right]}$$

1) Monod function for growth:

$$\frac{1}{\dot{N}} \left(\frac{d\dot{N}}{dt}\right) = \frac{\mu_{max} S}{(K_s + S)}$$

2) Relationship between growth and substrate removal:

$$\frac{d\dot{N}}{dt} = -Y_n \left(\frac{dS}{dt}\right)$$

3) Monod function for substrate removal:

$$\frac{dS}{dt} = -\frac{1}{Y_n} \left(\frac{d\dot{N}}{dt}\right) = -\frac{\dot{N}}{Y_n} \left(\frac{\mu_{max} S}{K_s + S}\right)$$

4) Formulate the derivative of the inverse of the Monod function for substrate removal:

$$\frac{dt}{dS} = -\frac{Y_n}{\dot{N}} \left(\frac{K_s + S}{\mu_{max} S}\right) = -\frac{Y_n}{\dot{N} \mu_{max}} \left[\left(\frac{K_s}{S}\right) + 1\right]$$

5) Integrate dt/dS between $S = S_i$ and $S = S_1$ to obtain the treatment time required to transform S_i into S_1:

$$t_n = \frac{Y_n}{\mu_{max} \dot{N}} \left[S_i - S_1 + K_s \ln \left(\frac{S_i}{S_1}\right)\right]$$

6) Express t_n in terms of μ_{na}, $\dot{N}$, Y_n, and $S_i - S_1$ per Equation 23:

$$t_n = \frac{Y_n\ (S_i - S_1)}{\mu_{na}\ \dot{N}}$$

7) Combine Steps 5 and 6, eliminating t_n, Y_n, and $\dot{N}$:

$$\frac{S_i - S_1}{\mu_{na}} = \frac{1}{\mu_{max}} \left[S_i - S_1 + K_s \ln \left(\frac{S_i}{S_1} \right) \right]$$

8) Solve for μ_{na}:

$$\mu_{na} = \frac{\mu_{max}\ (S_i - S_1)}{\left[S_i - S_1 + K_s \ln \left(\frac{S_i}{S_1} \right) \right]}$$

9) Eliminate S_i:

$$S_i = \frac{S_0 + r\ S_1}{1 + r}$$

$$S_i - S_1 = \frac{S_0 - S_1}{1 + r}$$

$$\mu_{na} = \frac{\mu_{max}\ (S_0 - S_1)}{S_0 - S_1 + (1 + r)\ K_s \ln \left[\frac{S_0 + r\ S_1}{(1 + r)\ S_1} \right]}$$

10) Consider that $\mu_{na} = \mu_{nr}$ for PF reactors per Equation 26.

APPENDIX B

DATA FROM AERATION TRAIN PROFILE SAMPLINGS

TABLE B-1. AERATION TRAIN PROFILE DATA FOR FLOW SHEET I (11/14/73)

Sampling Point		(1) Diss. TKN (mg/ℓ)	(1) NH_4^+-N (mg/ℓ)	(1) $NO_2^-+NO_3^-$ -N (mg/ℓ)	(1) Diss. Ortho-Phosphorus (mg/ℓ)	(1) Diss. COD (mg/ℓ)	(1) Diss. TOC (mg/ℓ)	(2) ML Settled Volume (ml)	(3) SS After 30 min Settling
Primary Settler Effluent				0.1	4.1				
Cell 1	Anoxic	15.8		0.1	1.0	111	36	500	10
Cell 2	Aerobic	10.4	9.2	1.2	0.4	32	20	400	6
Cell 3	Aerobic	9.6	8.1	3.6	0.2	24	22	400	4
Cell 4	Aerobic	7.5	6.3	5.5	0.2	24	20	350	4
Cell 5	Aerobic	5.7	4.6	7.2	0.1	48	20	350	5
Cell 6	Aerobic	4.1	3.0	8.7	0.1	24	16	400	4
Cell 7/8	Anoxic	3.9	2.7	4.9	0.1	16	18	350	5
Cell 9	Anoxic	4.4	3.1	3.8	0.1	16	20	350	5
Cell 10	Anoxic	4.1	2.9	2.1	0.1	16	18	400	4
Cell 11	Anoxic	3.7	2.8	0.9	0.2	24	18	350	3
Cell 12	Aerobic	2.2	1.2	3.1	0.1	40	20	350	3
Chlorine Contact Chamber Effluent			0.1	1.8	0.2				

(1) ML supernatant, filtered.
(2) After 30 minutes settling.
(3) In ML supernatant.

Additional ML Data:
MLSS = 3300 mg/ℓ; MLVSS = 2200 mg/ℓ;
Flow-Through-Time = 29 min per cell (except Cell 7/8)
Started to sample cells at 8:50 AM.

TABLE B-2. AERATION TRAIN PROFILE DATA FOR FLOW SHEET I (11/27/73)

Sampling Point		(1) pH	(1) Total Alk. (mg/ℓ as $CaCO_3$)	(1) Diss. TKN (mg/ℓ)	(1) NH_4^+-N (mg/ℓ)	(1) $NO_2^- + NO_3^-$ -N (mg/ℓ)	(1) Diss. Ortho-phosphorus (mg/ℓ)	(1) Diss. COD (mg/ℓ)	(1) Diss. TOC (mg/ℓ)	(2) ML Settled Volume (ml)	(3) SS After 30 min Settling
Primary Settler Effluent				22.1	14.5	0.1	4.4		112		
Cell 1	Anoxic	7.5	162	14.9	11.2	0.1	8.4	65+	39	460	20
Cell 2	Aerobic	7.6	146	9.0	7.5	1.6	0.67	65	27	380	12
Cell 3	Aerobic	7.5	124	4.8	4.1	5.1	0.17	24	23	390	7
Cell 4	Aerobic	7.4	108	2.5	1.6	7.8	0.08	8	20.5	370	7
Cell 5	Aerobic	7.4	98	1.7	0.3	9.7	0.05	24	N.AV.	350	8
Cell 6	Aerobic	7.4	96	1.1	0.2	10.1	0.04	16	19	340	6
Cell 7/8	Anoxic	7.5	112	1.1	0.2	5.8	0.03	8	21	340	11
Cell 9	Anoxic	7.5	112	1.1	0.2	5.7	0.02	24	21.5	340	10
Cell 10	Anoxic	7.6	118	1.4	0.3	4.1	0.02	33	22.5	390	9
Cell 11	Anoxic	7.4	116	1.1	0.4	2.3	0.04	16	20	370	11
Cell 12	Aerobic	7.5	122	1.1	0.1	3.1	0.03	16	20	350	12
Final Settler				1.0	0.2	1.7	0.03				

+ Doubtful values.
(1) ML supernatant, filtered.
(2) After 30 minutes settling.
(3) In ML supernatant.

Additional ML Data:
MLSS = 3600 mg/ℓ; MLVSS = 2200 mg/ℓ;
Flow-Through-Time = 29 min per cell (except Cell 7/8);
Started to sample cells at 8:50 AM.

TABLE B-3. AERATION TRAIN PROFILE DATA FOR FLOW SHEET I (12/5/73)

Sampling Point		(1) pH	(1) Total Alk. (mg/ℓ as $CaCO_3$)	(1) Diss. TKN (mg/ℓ)	(1) NH_4^+-N (mg/ℓ)	(1) NO_2^-+NO_3^- -N (mg/ℓ)	(1) Diss. Ortho-Phosphorus (mg/ℓ)	(1) Diss. COD (mg/ℓ)	(1) Diss. TOC (mg/ℓ)	(2) ML Settled Volume (ml)	(3) SS After 30 min Settling
Primary Settler Effluent				25.8	9.9	0.1	2.2		100		
Cell 1	Anoxic	7.5	164	15.7	11.3	0.1	7.6	87	47.0	440	30
Cell 2	Aerobic	7.6	134	10.4	8.1	2.2	1.10	40	29.5	480	10
Cell 3	Aerobic	7.6	120	6.4	4.6	4.9	0.27	24	23.5	560	45
Cell 4	Aerobic	7.5	100	3.6	1.8	8.4	0.21	32	18.5	360	5
Cell 5	Aerobic	7.5	94	1.7	0.4	9.7	0.16	32	18.0	430	1
Cell 6	Aerobic	7.6	88	2.0	0.2	10.6	0.14	16	16.9	350	1
Cell 7/8	Anoxic	7.5	100	1.4	0.2	6.7	0.10	24	14.0	360	7
Cell 9	Anoxic	7.5	108	1.7	0.3	5.9	0.13	32	22.5	400	9
Cell 10	Anoxic	7.7	110	1.4	0.3	4.8	0.11	32	21	360	10
Cell 11	Anoxic	7.6	116	1.4	0.4	2.6	0.16	32	23	400	4
Cell 12	Aerobic	7.5	110	1.4	0.1	3.3	0.11	32	23	380	4
Final Settler				1.1	0.2	2.5	0.02	24	20		
Chlorine Contact Chamber				0.9	0.1	2.2	0.03	24	18.5		

(1) ML supernatant, filtered.
(2) After 30 minutes settling.
(3) In ML supernatant.

Additional ML Data:
MLSS = 4000 mg/ℓ; MLVSS = 2500 mg/ℓ;
Flow-Through-Time = 29 min per cell (except Cell 7/8);
Sampling started at 8:00 AM (Prim. Settler).

TABLE B-4. AERATION TRAIN PROFILE DATA FOR FLOW SHEET IV (9/18/74)

Sampling Point		Dissolved TKN (mg/ℓ)	NH_4^+-N (mg/ℓ)	NO_2^-+NO_3^- -N (mg/ℓ)	Dissolved Orthophosphorus (mg/ℓ)
Surge Tank Effluent		15.4	11.4	0.1	4.6
Cell 1	Aerobic	8.12	6.6	0.2	0.2
Cell 2	Aerobic	5.9	4.5	2.2	0.08
Cell 3	Aerobic	4.2	2.7	4.2	0.05
Cell 4	Aerobic	3.1	1.9	4.7	0.04
Cell 5	Anoxic	3.1	2.0	3.5	0.03
Cell 6	Anoxic	3.1	2.1	1.2	0.03
Cell 7/8	Anoxic	3.6	2.4	0.1	0.06
Cell 9	Anoxic	3.6	2.6	0.1	0.05
Cell 10	Anoxic	3.6	2.7	0.1	0.04
Cell 11	Aerobic	1.7	0.8	1.7	0.03
Cell 12	Aerobic	1.4	0.2	2.5	0.05

Additional Primary Effl. Data:
TKN = 26.6 mg/ℓ
Total P = 12.4 mg/ℓ
Dissolved P = 5.8 mg/ℓ

Additional ML Data:
MLSS = 2800 mg/ℓ
MLVSS = 1800 mg/ℓ

ML Flow-Through-Time = 26 min per cell (except Cell 7/8).

TABLE B-5. AERATION TRAIN PROFILE DATA FOR FLOW SHEET IV (10/10/74)

Sampling Point		Sampling Time	DO (mg/ℓ)	Temp. (C)	pH	Total Alk. (mg/ℓ as $CaCO_3$)	Dissolved TKN (mg/ℓ)	NH_4^+-N (mg/ℓ)	$NO_2^-+NO_3^-$ -N (mg/ℓ)	Dissolved Orthophosphorus (mg/ℓ)
Surge Tank Effl.		8:39			7.5	173	17.08	12.0	0.2	2.2
Cell 1	Aerobic	9:06	3.2	23.5	7.2	135	12.32	9.4	2.4	0.77
Cell 2	Aerobic	9:33	3.9	23.4	7.2	136	10.08	8.5	4.0	0.42
Cell 3	Aerobic	10:00	3.4	23.2	7.1	117	7.28	5.5	5.0	0.25
Cell 4	Aerobic	10:27	4.1	23.0	7.0	118	5.60	4.1	7.1	0.24
Cell 5	Anoxic	10:54	0.1	23.0	7.1	122	5.60	4.3	4.8	0.21
Cell 6	Anoxic	11:21	0.1	23.0	7.1	129	5.60	4.6	2.7	0.20
Cell 7/8	Anoxic	12:13	0.1	22.5	7.1	143	6.44	5.1	0.1	0.49
Cell 9	Anoxic	12:39	0.1	22.5	7.3	142	5.88	5.0	0.1	0.25
Cell 10	Anoxic	1:05	0.1	22.5	7.2	144	5.88	5.0	0.1	1.08
Cell 11	Aerobic	1:31	5.8	22.1	7.3	124	3.64	2.4	2.4	0.33
Cell 12	Aerobic	1:57	6.5	22.2	7.2	109	1.96	0.4	4.1	0.15

Additional Surge Tank Effluent Data: Total P = 9.0 mg/ℓ; Dissolved P = 2.3 mg/ℓ; TKN = 30.2 mg/ℓ.
Additonal ML Data: MLSS = 3930 mg/ℓ; MLVSS = 2600 mg/ℓ; Flow-Through-Time = 26.5 min per cell (except Cell 7/8).
Fractional Rate of Return Sludge Recirculation = 0.44.

TABLE B-6. AERATION TRAIN PROFILE DATA FOR FLOW SHEET IV (11/14/74)

Sampling Point		Sampling Time	DO (mg/ℓ)	Temp. (C)	pH	Total Alk. (mg/ℓ as $CaCO_3$)	Dissolved TKN (mg/ℓ)	NH_4^+-N (mg/ℓ)	$NO_2^-+NO_3^-$ -N (mg/ℓ)	Dissolved Orthophosphorus (mg/ℓ)
Surge Tank Effl.		8:41			7.3	199	17.9	11.4	0.1	3.7
Cell 1	Aerobic	9:05	0.6	19	7.6	193	13.4	9.9	0.1	3.0
Cell 2	Aerobic	9:30	2.5	19	7.7	187	11.2	9.0	1.0	2.0
Cell 3	Aerobic	9:55	5.6	18.8	7.7	173	8.7	7.4	2.5	1.4
Cell 4	Aerobic	10:19	6.7	18.8	7.8	169	7.6	5.9	3.8	1.0
Cell 5	Anoxic	10:44	0.1	18.9	7.6	168	6.4	5.6	3.1	0.78
Cell 6	Anoxic	11:09	0.1	18.6	7.7	166	6.6	5.4	1.2	0.76
Cell 7/8	Anoxic	11:58	0.1	18.6	7.6	171	6.4	5.7	0.1	2.3
Cell 9	Anoxic	12:23	0.1	18.6	7.5	169	6.9	5.7	0.1	3.2
Cell 10	Anoxic	12:47	0.1	18.5	7.7	177	6.2	5.6	0.1	3.8
Cell 11	Aerobic	1:12	4.8	18.5	7.7	165	5.0	4.1	1.0	1.9
Cell 12	Aerobic	1:37	7.4	18.4	7.7	156	3.4	2.8	2.2	1.4

Additional Surge Tank Effluent Data: Total P = 7.62 mg/ℓ; Soluble P = 3.52 mg/ℓ; TKN = 25.5 mg/ℓ.
Additional ML Data: MLSS = 2670 mg/ℓ; MLVSS = 1580 mg/ℓ; Flow-Through-Time = 25 min per cell (except Cell 7/8).
Fractional Rate of Return Sludge Recirculation = 0.42.

TABLE B-7. AERATION TRAIN PROFILE DATA FOR FLOW SHEET VI (6/12/75)

Sampling Point	Sampling Time	DO (mg/ℓ)	pH	Total Alk. (mg/ℓ as $CaCO_3$)	Dissolved TKN (mg/ℓ)	NH_4^+-N (mg/ℓ)	$NO_2^-+NO_3^-$-N (mg/ℓ)	Dissolved Ortho-Phosphorus (mg/ℓ)	Dissolved TOC (mg/ℓ)
Prim. Settler*	9:00	0.3	7.45	272	29.7	26.0	0.1	6.6	84.5
Mixing Chamber**	9:00	0.7	7.35	205	24.6	20.5	0.1	9.3	77.5
Cell 1 Aerobic	9:28	1.8	7.5	172	14.0	12.8	2.7	5.0	15.0
Cell 2 Aerobic	9:57	2.05	7.4	127	7.6	5.8	10.1	4.9	11.5
Cell 3 Aerobic	10:25	3.2	7.3	94	2.1	0.7	15.5	4.6	8.5
Cell 4 Anoxic	10:54	0.0	7.2	102	2.5	1.0	13.5	4.3	9.0
Cell 5 Anoxic	11:23	0.0	7.5	111	2.6	1.4	11.6	4.1	8.5
Cell 6 Anoxic	11:51	0.0	7.6	116	3.0	1.6	9.7	4.0	8.0
Cell 7/8 Anoxic	12:45	0.0	7.6	132	3.6	2.3	5.1	3.9	9.5
Cell 9 Anoxic	1:13	0.0	7.7	136	3.9	2.6	4.3	3.9	9.0
Cell 10 Anoxic	1:40	0.0	7.7	142	3.8	2.8	3.3	3.8	9.0
Cell 11 Aerobic	2:07	3.7	7.7	124	1.5	0.1	6.4	3.7	8.0
Cell 12 Aerobic	2:34	7.5	7.6	123	1.3	0.1	6.8	3.5	9.5

* Effluent Trough of Primary Settler.

** Admixing Return Sludge to Primary Effluent.

Temperature of Aeration Cells: first = 28.5 C; last = 25 C.

Additional Primary Effluent Data: TKN = 37 mg/ℓ; Total P = 9.5 mg/ℓ.

Additional ML Data: MLSS = 4150 mg/ℓ; MLVSS = 2600 mg/ℓ; Flow-Through-Time = 28 min per cell (except Cell 7/8).